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A MANUAL OF ELECTROSTATIC PRECIPITATOR TECHNOLOGY

PART II - APPLICATION AREAS

SOUTHERN RESEARCH INSTITUTE
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PART II - APPLICATION AREAS

Prepared Under Contract CPA 22-69-73

for

THE NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
Division of Process Control Engineering
Cincinnati, Ohio

by

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Engineering Research and
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and

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Southern Research Institute
Birmingham, Alabama
August 25, 1970
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FOREWORD

This manual of electrostatic precipitator technology was prepared under Contract CPA 22-69-73 for the National Air Pollution Control Administration, Division of Process Control Engineering, with Mr. Timothy W. Devitt serving as Project Officer. The manual is presented in two parts. Part I covers the fundamentals of electrostatic precipitation and Part II covers the application of electrostatic precipitators to control of emissions in various application areas.

Part I of this manual was prepared on the basis of a comprehensive review of precipitator literature and discussions with those active in research in the field of electrostatic precipitation. Part II of the manual was prepared on the basis of information prepared by Research-Cottrell, Inc., Bound Brook, New Jersey, based on records covering installations in the various application areas. These installations include some designed by other precipitator manufacturers, as well as those designed by Research-Cottrell.

To supplement the data furnished by Research-Cottrell, Southern Research Institute conducted a survey of a number of installations in various industries. In some instances, considerable data were available covering both design and performance of the precipitators. In other instances, data were not available. Consequently, the extent and quality of the information covering the application areas varies.

Rust Engineering Company, Division of Litton Industries, Birmingham, Alabama, furnished data on the application of electrostatic precipitators in the pulp and paper industry.

Dr. Harry J. White, Head of the Department of Applied Science, Portland State University, served as consultant on the program and contributed heavily in the review of precipitator fundamentals.

ABSTRACT

This manual is a comprehensive report on the state of technology of electrostatic precipitation for particulate emission control. It is intended to serve as a basic handbook on the subject of electrostatic precipitation covering the entire system including fundamental theory, design, operation, and application.

The purpose of the manual is to provide a source of information that will be of benefit to the researcher, manufacturer, and user of electrostatic precipitators. The information and data are based on an electrostatic precipitator systems study performed under Contract No. CPA 22-69-73 for the National Air Pollution Control Administration.

The systems study included a review of literature pertaining to electrostatic precipitation, a comprehensive review of precipitator fundamentals, a survey and review of the use of precipitators in each of eight major application areas, and a review of the potential for electrostatic precipitators in new applications.

In order to provide information useful to a wide range of users, this manual is organized to present a general review of the precipitation process, with qualitative relationships between the various precipitator functions, and a detailed quantitative analysis of each of the system functions. Particular emphasis is placed on the validity of the theoretical derivations in view of the assumptions made and the significance of the interaction of variables of importance in precipitator operation.

Further discussions include mechanical design considerations, energization equipment, gas flow, operational problems, and the current utilization of precipitators in the major application areas.

TABLE OF CONTENTS
PART I - FUNDAMENTALS

Page No.

DESCRIPTIVE SUMMARY 1

Corona Generation 1
Particle Charging 6
Particle Collection 9
Removal 11
Electrical Energization 12
Systems Analysis 13
Design 14
Mechanical Components .... 1[illegible]
Gas Flow 17
Resistivity 18
Measurement of Performance 19
Troubleshooting and Maintenance 20
Electrostatic Augmentation and Unusual Designs 21

CHAPTER 1. CORONA GENERATION 23

1.1 Current Voltage Relationship 30
1.2 Temperature and Pressure Effects 34
1.3 Practical Aspects of Corona Theory on Electrostatic Precipitation 40
1.4 Practical Aspects of Corona Generation 53
Bibliography 55

CHAPTER 2. THE ELECTRIC FIELD 57

2.1 Space Charge Phenomena 57
2.2 Practical Aspects of the Electric Field 66
Bibliography 68

CHAPTER 3. PARTICLE CHARGING 69

3.1 Field Charging 70
3.2 Diffusion Charging 78

Page No.

3.3 Combination of Field and Diffusion Charging......... 81
3.4 Practical Aspects of Particle Charging 85
Bibliography 95

CHAPTER 4. PARTICLE COLLECTION 96

4.1 Particle Kinetics 96
4.2 Particle Collection with Laminar Gas Flow.......... 103
4.3 Particle Collection with Turbulent Gas Flow......... 105
4.4 Factors Modifying the Particle Collection in a Precipitator................................ 114
4.5 Practical Aspects of Particle Collection 117
4.6 Recent Modifications to the Deutsch Equation 118
Bibliography 123

CHAPTER 5. GAS FLOW.................................. 124

5.1 Introduction.. 124
5.2 Straighteners 138
5.3 Splitters... 139
5.4 Transformation Splitters 139
5.5 Vanes 141
5.6 Diffusion Plates 144
5.7 Modelling... 146
Bibliography 151

CHAPTER 6. RAPPING AND REENTRAINMENT 152

6.1 Rapping 152
6.2 Reentrainment 158
Bibliography 165

CHAPTER 7. RESISTIVITY AND CONDITIONING 166

7.1 Volume Conduction 166
7.2 Surface Conduction 167
7.3 Mechanisms of Surface Conditioning................. 171
7.4 Field Conditioning Studies 177
7.5 Moisture Conditioning............................... 177
7.6 Chemical Conditioning 179
Bibliography 186

Page No.

CHAPTER 8. ELECTROSTATIC PRECIPITATOR SYSTEMS ANALYSIS 187

8.1 Generalized Precipitator Model 188
8.2 Simplified System Model 190
8.3 Example Utilizing Existing Simplified System Model . 193
8.4 Limitations of Existing Model 199
8.5 Model Verification 200

CHAPTER 9. DESIGN METHODOLOGY 203

9.1 Design Method I 203
9.2 Design Method II 214
Bibliography 217

CHAPTER 10. MECHANICAL COMPONENTS 218

10.1 Tubular Type Precipitators 218
10.2 Plate Type Precipitators 218
10.3 Discharge Electrodes 221
10.4 Discharge Electrode Support 227
10.5 Collecting Electrodes 229
10.6 Electrode Cleaning Equipment 233
10.7 Hopper Systems 239
10.8 Dust Removal Systems 239
Bibliography 241

CHAPTER 11. POWER SUPPLIES AND CONTROLS 242

11.1 The Power System 242
Bibliography 265

CHAPTER 12. MEASUREMENTS 266

12.1 Measurement of Gas Flow and Efficiency 266
12.2 Velocity and Gas Volume 266
12.3 Determination of Dust Concentration 268
12.4 Computation 272
12.5 Sources of Error 276
12.6 Measurement of Resistivity 288
Bibliography 297

Page No.

CHAPTER 13. TROUBLESHOOTING AND MAINTENANCE...... 298

13.1 Gas Flow.. 301
13.2 Rapping.. 302
13.3 Hoppers and Ash Removal.......................... 304
13.4 Troubleshooting Techniques....................... 304
13.5 Maintenance...................................... 305

CHAPTER 14. ELECTROSTATIC AUGMENTATION AND UNUSUAL PRECIPITATION DESIGNS........... 311

14.1 Wet Scrubbers.................................... 311
14.2 Electrostatically Augmented Packed Bed Filters..... 314
14.3 Fabric Filter.................................... 315
14.4 Mechanical Collection............................ 315
14.5 Space Charge Precipitation....................... 316
Bibliography...................................... 322

TABLE OF CONTENTS
PART II - APPLICATION AREAS

Page No.

DESCRIPTIVE SUMMARY 324

- Electric Power Generation 324
- Pulp and Paper 328
- Iron and Steel Industry 330
- Rock Products 337
- Chemical Industry 340
- Municipal Incinerators 343
- Petroleum Industry 344
- Nonferrous Metals 345
- High Temperature, High Pressure Gas Cleaning 346
- New Application Areas 347

CHAPTER 15. ELECTROSTATIC PRECIPITATORS IN THE ELECTRIC UTILITY INDUSTRY 348

- 15.1 Fly Ash Characteristics 351
- 15.2 Flue Gas Characteristics 360
- 15.3 Historical Development of Fly Ash Precipitators 363
- 15.4 Design of Fly Ash Precipitators 365
- 15.5 Summary of Precipitator Operating Conditions 383
- 15.6 Precipitator Installations and Economics 393
- 15.7 Trends 408
- Bibliography 416

CHAPTER 16. ELECTROSTATIC PRECIPITATORS IN THE PULP AND PAPER INDUSTRY 418

- 16.1 Process Descriptions 421
- 16.2 Character of Recovery Boiler Particulate Emission 427
- 16.3 Character of Sulfate Process Flue Gases 432
- 16.4 Historical Developments of Electrostatic Precipitators for Recovery Furnaces 435
- 16.5 Precipitator Applications 445
- 16.6 Summary of Precipitator Operating Conditions 455

Page No.

16.7 Assessment of the State of the Art 462
16.8 General Observations and Discussion of Trends.... 467
Bibliography 473

CHAPTER 17. ELECTROSTATIC PRECIPITATORS IN THE IRON AND STEEL INDUSTRY................ 474

17.1 Introduction 474
17.2 Application to Coke Ovens...................... 476
17.3 Application to Sinter Plants..................... 487
17.4 Application to Blast Furnaces.................... 502
17.5 Application to Open Hearth Furnaces 529
17.6 Application to Basic Oxygen Converters........... 546
17.7 Application to Electric Arc Furnaces 567
17.8 Application to Scarfing Machines................. 584
17.9 Application to Iron Cupolas 592
Bibliography 604

CHAPTER 18. ELECTROSTATIC PRECIPITATORS IN THE ROCK PRODUCTS INDUSTRY................ 609

18.1 Manufacture of Portland Cement 609
18.2 Character of Emissions from Cement Kilns 615
18.3 Electrostatic Precipitators for Control of Cement Dusts 621
18.4 Electrostatic Precipitators for the Gypsum Industry 640
Bibliography 659

CHAPTER 19. ELECTROSTATIC PRECIPITATORS IN THE CHEMICAL INDUSTRY..................... 660

19.1 Major Fields of Application 660
19.2 Sulfuric Acid Production Processes 660
19.3 Design and Operating Parameters for Sulfuric Acid Mist Precipitators........................... 666
19.4 Sulfuric Acid Mist Precipitator Installations and Cost Data 675
19.5 Precipitators for the Elemental Phosphorus Industry 684
19.6 Precipitators for the Phosphoric Acid Industry 704
19.7 Precipitators for the Carbon Black Industry....... 722
Bibliography 728

Page No.

CHAPTER 20. THE APPLICATION OF ELECTROSTATIC PRECIPITATORS IN CLEANING MUNICIPAL INCINERATOR DUSTS........................ 729

20.1 Types of Incinerators 729
20.2 Incinerator Capacities 730
20.3 Refuse Properties 731
20.4 Emission Properties 732
20.5 Incinerator Emission Control 737
20.6 Use of Electrostatic Precipitators on Municipal Incinerators........................... 740
20.7 Economics 752
Bibliography 754

CHAPTER 21. ELECTROSTATIC PRECIPITATORS IN THE PETROLEUM INDUSTRY........................... 755

21.1 Introduction........................... 755
21.2 Catalytic Cracking........................... 756
21.3 Detarring........................... 761
21.4 Development of Electrostatic Precipitators in the Petrochemical Industry........................... 768
Bibliography........................... 785

CHAPTER 22. THE APPLICATION OF ELECTROSTATIC PRECIPITATORS IN THE NONFERROUS METALS INDUSTRY 786

22.1 Historical Development 786
22.2 Nonferrous Metal Processing 788
22.3 Electrolytic Reduction of Aluminum........................... 798
22.4 Production of Primary Copper 799
22.5 Production of Primary Lead 810
22.6 Zinc Reduction 818
22.7 Summary of Precipitator Operation Conditions 822
Bibliography 836

Page No.

CHAPTER 23. THE APPLICATION OF ELECTROSTATIC PRECIPITATORS TO THE CLEANING OF HIGH PRESSURE, HIGH TEMPERATURE GASES..... 837

23.1 High Pressure, High Temperature Processes 837
23.2 Influence of Temperature and Pressure on Precipitator Performance 838
23.3 Precipitators for High Temperature, High Pressure Operation.. 840
23.4 Cleaning of Natural Gas in Pipelines.............. 841
Bibliography 844

CHAPTER 24. NEW PRECIPITATOR APPLICATIONS 845

24.1 Application of Electrostatic Precipitators 856
Bibliography 875

LIST OF FIGURES
PART II

Figure No.		Page No.
15.1	Frequency Distribution of Percent Ash in Coal Burned in 48 different Power Plant Boilers on Which Precipitators are Installed...............	353
15.2	Typical Fly Ash Particle Size Distribution for Pulverized Fuel Furnaces	354
15.3	Typical Range of Particle Size Distributions for Fly Ash from Cyclone Furnaces................	356
15.4	Trends in Resistivity of Fly Ash with Variations in Flue Gas Temperature and Coal Sulfur Content.....................................	361
15.5	Frequency Distribution of Percent Total Sulfur in Coal Burned in 48 Different Power Plant Boilers on Which Precipitators are Installed.....	362
15.6	Distribution of Precipitation Rates for Fly Ash Precipitators............................	366
15.7	Relationship Between Collection Efficiency and Collecting Surface Area to Gas Flow Ratio for Various Coal Sulfur Contents..................	369
15.8	Variation in Precipitation Rate Parameter With Sulfur Content of the Coal......................	370
15.9	Relationship Between Precipitation Rate Parameter and In-situ Resistivity........................	371
15.10	Relationship Between Collection Efficiency and Corona Power for Fly Ash Precipitators	373
15.11	Relationship Between Precipitation Rate Parameters and Corona Power Density in Electric Utility Installations................................	375
15.12	Variation in Efficiency with Degree of Sectionalization.............................	376
15.13	Distribution of Design Precipitator Electrical Field Strength...............................	378
15.14	Cold End Temperature and Material Selection Guide..	381
15.15	Distribution of Precipitator Inlet Dust Loading.......................................	384
15.16	Distribution of Precipitator Inlet Gas Temperature................................	385

Figure No.		Page No.
15.17	Distribution of Design Precipitator Input Power..	386
15.18	Distribution of Precipitator Gas Velocity........	387
15.19	Comparison Between Measured Precipitation Rate Parameters and Values Calculated by Regression Analysis..........................	390
15.20	Comparison Between Measured Precipitation Rate Parameter and Value Calculated from Regression Analysis (SRI).....................	391
15.21	Comparison of Gas Volume with Electrostatic Precipitator Cleaning to Total Gas Volume Calculated from Coal Burned	397
15.22	Percentage of Flue Gas from Coal-Firing with Electrostatic Precipitator Cleaning.............	398
15.23	Average Yearly Precipitator Gas Volume for the 5-Year Periods from 1923 through 1969.........	399
15.24	Average Yearly Precipitator Design Efficiency, Percent (Prorated on a cfm Basis) Over the Period 1923 through 1969......................	400
15.25	Electric Utility Electrostatic Precipitator Costs Vs. Volume Flow for Collection Efficiencies Between 90 and 95%..........................	403
15.26	Electric Utility Electrostatic Precipitator Costs Vs. Volume Flow for Collection Efficiencies Between 95 and 99%..........................	404
15.27	Electric Utility Electrostatic Precipitator Costs Vs. Volume Flow for Collection Efficiencies Greater than 99%...........................	405
15.28	Precipitator Costs (Average) Over the Period 1923 through 1969	407
16.1	United States Production of Chemical and Semi-Chemical Pulps (1900-1965)..............	422
16.2	Simplified Flow Sheet, Kraft Pulping Process ...	425
16.3	Photomicrograph of Particles Collected from a Pulp Mill Recovery Unit (Temperature, Approximately 350°F).........................	428
16.4	Photomicrograph of Particles Collected from a Pulp Mill Recovery Unit (Temperature, Approximately 280°F).........................	428

Figure No. Page No.

16.5 Photomicrograph of Particles Resulting from Mixing of Gases and Dust 429
16.6 Count Distribution of the Sample in Figure 16.3 429
16.7 Particle Size Distribution of Recovery Furnace Fume 430
16.8 Electrical Resistivity of Salt Cake (Laboratory Measurement with 1-2% Moisture in Gas) 433
16.9 Electrical Resistivity of Sodium Sulfate as a Function of Moisture in Gas at 300°F. 434
16.10 Vertical Flow Salt Cake Precipitator With Dry (Hopper) Dust Removal System................. 436
16.11 Typical Dry Bottom Electrostatic Precipitator Recovery Boiler Installation.................. 437
16.12 Vertical Flow Salt Cake Precipitator With Wet Bottom Dust Removal System 438
16.13 Typical Wet Bottom Electrostatic Precipitator Recovery Boiler Installation.................. 439
16.14 Horizontal Flow Salt Cake Precipitator With Wet Bottom Dust Removal System 441
16.15 Typical Venturi Scrubber Recovery System Installation 443
16.16 Recovery Boiler Installation With Cyclone Evaporation and Wet Bottom Electrostatic Precipitator 444
16.17 Estimate of Total Sulfate Pulp Production With Electrostatic Precipitators Installed on Recovery Furnace Gases.............................. 446
16.18 Installed Precipitator Gas Volume Over the Period 1939 through 1969...................... 448
16.19 FOB Costs for Recovery Boiler Precipitator 449
16.20 Erected Costs of Recovery Boiler Precipitators - 1969 Dollars, 99+% Efficiency................. 450
16.21 Erected Costs for Recovery Boiler Precipitators - 1969 Dollars, 90-99% Efficiency............... 451
16.22 Electrostatic Precipitator Costs (1960-1969) 454
16.23 Collection Efficiency Vs. Input Power per Thousand cfm in Pulp and Paper Installations.... 456

Figure No.		Page No.
16.24	Distribution of Precipitator Gas Velocity	457
16.25	Distribution of Precipitator Inlet Dust Loading	458
16.26	Distribution of Precipitator Connected Input Power	459
16.27	Distribution of Precipitator Inlet Gas Temperature	460
16.28	Distribution of Precipitator Design Electrical Field Strength	461
16.29	Histogram of Precipitation Rate Parameters for 15 Pulp and Paper Installations	464
16.30	Relationship Between Precipitation Rate Parameter and Power Density	464
16.31	Relationship Between Design and Test Collection Efficiencies Vs. A/V Ratio for 11 Electrostatic Precipitators for Pulp and Paper Installations	466
16.32	Accumulative Distribution of Precipitator Gas Volume (1939-1969)	469
16.33	Comparison of Actual Performance to Design Performance - Basis: Ratio of Migration Velocities Calculated Using Deutsch Equation	470
17.1	A Flow Chart Showing the Steps in the Steelmaking Process from the Basic Raw Material and Scrap Input to the Finished Product	475
17.2	Rates of U. S. Steel Production by the Various Processes and the Total Steel Production for the Period 1920 to 1969, with Projection for the Period 1969-1980	477
17.3	The Production of Steel, Pig Iron, and Steel Scrap Used During the Period 1920-1969	478
17.4	Equipment Used to Remove the Particulate and Heavy Tar from Coke Oven Gas Prior to Recovery of Various Hydrocarbons	481
17.5	A Typical Pipe-Type Precipitator for the Collection of Tar	483
17.6	An Integral Tar Pipe-Type Precipitator	484
17.7	Concentric Ring Detarrer	485

Figure No.		Page No.
17.8	Iron Ore Sintering Machine Gas Cleaning With Electrostatic Precipitator.....................	491
17.9	Photomicrograph of Emissions from Sinter Plants, 7800X, after Mechanical Collector......	493
17.10	Particle Size Distribution by Weight of Sintering Machine Dust................................	494
17.11	Electrical Resistivity of Sintering Machine Dust...	495
17.12	Installed Capacity of Electrostatic Precipitators for Sinter Machine Dust Collection for the Period 1951-1969.................................	498
17.13	Variation in Collection Efficiency of Sinter Machine Precipitators with Collection Surface Area to Gas Flow Ratio.......................	500
17.14	Range of Installed Costs for Various Gas Volumes for Precipitators Operating on Sinter Machines..	501
17.15	Range of FOB Precipitator Costs for Various Gas Volumes	503
17.16	Flow Diagram for Wet Cleaning Iron Blast Furnace Gas with Electrostatic Precipitator	505
17.17	Dust Rate and Wind Rate as a Function of the Sinter Plus Taconite in the Blast Furnace Burden	508
17.18	Typical Wet Type Pipe Precipitator for Cleaning Blast Furnace Gases	514
17.19	Blast Furnace Statistics for Period 1920-1969...	516
17.20	Blast Furnace Precipitator Gas Flow Capacity for the Period 1931-1969......................	517
17.21	Distribution of Precipitator Inlet Gas Temperature	518
17.22	Distribution of Precipitator Inlet Dust Loading ..	519
17.23	Distribution of Precipitator Gas Velocity	520
17.24	Distribution of Precipitator Field Strength.......	521
17.25	Distribution of Precipitator Input Power........	522
17.26	Design Efficiency Trends Over the Period 1931-1969 Prorated on acfm Basis for Blast Furnace Installations........................	523

Figure No. Page No.

17.27 Relationship Between Collection Efficiency and Specific Collection Area for Electrostatic Precipitators Operating on Blast Furnaces...... 525

17.28 Range of FOB Costs as a Function of Collecting Surface Area for Blast Furnace Installations..... 526

17.29 Flow Diagram for Dry Cleaning Ferromanganese Blast Furnace Gas with Electrostatic Precipitator.................................. 528

17.30 Particle Size Analysis of Open Hearth Dust from a Furnace Utilizing an Oxygen Lance 538

17.31 Electrical Resistivity of Red Oxide Fume from Three Oxygen Lanced Open Hearth Furnaces..... 539

17.32 Flow Diagram for Cleaning Open Hearth Furnace Gas With an Electrostatic Precipitator.......... 541

17.33 Schematic Diagram Showing a Typical Electrostatic Precipitator Installation on an Open Hearth Furnace 543

17.34 Cumulative Installed Capacity of Electrostatic Precipitators for Open Hearth Installations...... 545

17.35 Relationship Between Collection Efficiency and Collection Electrode to Volume Flow Ratio 547

17.36 Relationship Between Installed Power per Thousand cfm and Collection Efficiency for 5 Open Hearth Installations 548

17.37 Relationship Between Costs and Gas Flow Rate for Electrostatic Precipitators for 8 Open Hearth Installations................................. 550

17.38 Carbon Dioxide from Basic Oxygen Furnaces After Complete Combustion.................... 552

17.39 Resistivity Vs. Gas Temperature for BOF Furnace Dust................................ 555

17.40 Basic Oxygen Furnace Gas Cleaning With Electrostatic Precipitator 557

17.41 Cumulative Gas Volume from BOF Controlled by Electrostatic Precipitators 561

17.42 Relationship Between Collection Efficiency and Collection Area to Volume Flow Rate for BOF Installation............................. 562

Figure No. | Page No.

17.43 Erected Costs for a Limited Number of BOF Installations.................................... 565

17.44 Flow Diagram for Cleaning of Electric Furnace Exhaust Gases............................. 568

17.45 Variation of Fume Burdens During Oxygen Lancing.. 578

17.46 Particle Size Distribution by Weight of Electric Furnace Fume.............................. 581

17.47 Flow Diagram Showing Use of Electrostatic Precipitator for Cleaning Scarfing Machine Gas.. 586

17.48 Installed Precipitator Gas Volume for Scarfing Machine 588

17.49 Flow Diagram of Hot Blast Cupola Furnace Gas Cleaning System.......................... 593

17.50 Flow Diagram of Cold Blast Cupola Furnace Gas Cleaning System.......................... 594

17.51a Particle Size Distributions from Cold Blast Cupola Furnaces............................. 597

17.51b Particle Size Distributions from Hot Blast Cupola Furnaces............................. 597

17.52 Dust Output Survey as a Function of Coke Rate and Melt Rate for Hot Blast Cupolas 599

17.53 Dust Content of Top Gas as a Function of Specific Blast Volume................................ 600

17.54 Particle Size Ranges for Dusts from Cold and Hot Blast Cupolas 601

17.55 Laboratory Resistivity Measurements for Dusts Collected from Two Cupolas for Various Moisture Contents and Temperatures.................... 603

18.1 Production of Portland Cement in the United States (1900-1969)................................. 610

18.2 Production Statistics of the Gypsum Industry (1935-1969) 611

18.3 Manufacture of Portland Cement................ 612

18.4 Typical Particle Size Range of Cement Dust 618

18.5 Relationship Between Resistivity, Temperature, and Moisture Content for Cement Kiln Dust...... 620

Figure No.		Page No.
18.6	High Temperature Psychrometric Chart.........	623
18.7	Installed Precipitator Gas (Cumulative) for the Cement Industry.............................	627
18.8	Design Efficiency Trend for Cement Plant Precipitators (1912-1969)......................	628
18.9	Relationship Between Collection Efficiency and Specific Collection Area for Cement Industries...	635
18.10	Efficiency Vs. Power Rate for Cement Plants....	638
18.11	FOB and Erected Costs for Precipitators in the Cement Industry for Collection Efficiencies from 98-99.9%....................................	641
18.12	Schematic Diagram of a Gypsum Plant with Electrostatic Precipitator.....................	644
18.13	Typical Particle Size Distribution of Gypsum Dust.................................	646
18.14	Typical Laboratory and Field Resistivities of Gypsum Dust.................................	649
18.15	Electrostatic Precipitators in the Gypsum Industry (Installed Capacity) Data from Tables 18.14 and 18.15..............................	653
18.16	Collection Efficiency Vs. Specific Collecting Surface for Gypsum Rock Dryer Plant Preceded by Mechanical Collector......................	655
19.1	Typical Flow Chart for Sulfur-Burning Contact Plant..	662
19.2	Applications for Wet and Dry Electrostatic Precipitators in Contact Sulfuric Acid Plants Using Sulfur-Bearing Ores.........................	664
19.3	Typical Sulfuric Acid Mist Pipe-Type Precipitator.................................	667
19.4	Distribution of Sulfuric Acid Mist Precipitator Gas Velocity, 1945-1969......................	670
19.5	Distribution of Sulfuric Acid Mist Precipitator Inlet Mist Loading, 1945-1969.................	671
19.6	Distribution of Sulfuric Acid Mist Precipitator Inlet Temperature, 1945-1969.................	672

Figure No. Page No.

19.7 Distribution of Sulfuric Acid Mist Precipitator Input Power, 1945-1969........................ 673
19.8 Distribution of Sulfuric Acid Mist Precipitator Field Strength, 1945-1969...................... 674
19.9 Cumulative Gas Flow Capacity of Sulfuric Acid Mist Precipitators, 1945-1969.................. 677
19.10 Yearly Average Gas Flow Ratings of Sulfuric Acid Precipitators.................................. 678
19.11 Average Size of Acid Mist Precipitators in Terms of Gas Flow Capacity......................... 679
19.12 Design Efficiency Trend for Sulfuric Acid Mist Precipitators Weighted on acfm Basis, 1945-1969.................................. 680
19.13 Installed Cost of Sulfuric Acid Mist Precipitators, 1965-1969.................................. 682
19.14 Elemental Phosphorus Production by the Electric Arc Furnace Process.......................... 688
19.15 Typical Electric Furnace Phosphorus Precipitator.................................. 689
19.16 Cumulative Gas Flow Capacity of Elemental Phosphorus Precipitators Since 1938............ 693
19.17 Average Gas Flow Ratings of Elemental Phosphorus Process Precipitators, 1938-1969 ... 694
19.18 Average Size of Elemental Phosphorus Process Precipitators Over Period 1938 to 1969.......... 695
19.19 Distribution of Elemental Phosphorus Process Precipitator Design Gas Velocity, 1938-1969..... 698
19.20 Distribution of Elemental Phosphorus Process Precipitator Design Inlet Gas Temperature, 1938-1969.................................. 699
19.21 Distribution of Elemental Phosphorus Precipitator Design Inlet Loading, 1938-1969 700
19.22 Trends in Design Velocity for Elemental Phosphorus Precipitator, 1938-1969 705
19.23 Flow Sheet for Phosphoric Acid Production by Phosphorus Combustion....................... 706
19.24 Cumulative Phosphoric Acid Mist Precipitator Capacity Since 1928 710

Figure No. Page No.

19.25 Annual Phosphoric Acid Mist Precipitator Capacity for Period 1927-1956 711
19.26 Average Phosphoric Acid Mist Precipitator Size Over Period 1927-1956 712
19.27 Distribution of Phosphoric Acid Mist Precipitator Design Velocity, 1927-1969 717
19.28 Distribution of Phosphoric Acid Mist Precipitator Design Temperature, 1927-1969 718
19.29 Distribution of Phosphoric Acid Mist Precipitator Actual Inlet Loading, 1927-1969 719
19.30 Flow Diagram of Oil Furnace Process for Carbon Black Showing Precipitator Use 723

20.1 Particle Size Distribution of Incinerator Emissions Prior to Conditioning 735
20.2 Electrical Resistivity of Municipal Incinerator Dust 736
20.3 General Layout of Electrostatic Gas Cleaning Plants for Municipal Incinerators 743
20.4 Relationship Between Collection Efficiency and Specific Collection Area for Municipal Incinerators 745
20.5 Relationship Between Collection Efficiency and Delivered Corona Power for Municipal Incinerators 746
20.6 Relationship Between the Power Density and Precipitation Rate Parameter for Electrostatic Precipitators Operating on Effluents from Municipal Incinerators 747
20.7 Variation in Precipitation Rate Parameter with Gas Temperature for Municipal Incinerator Precipitators 751

21.1 Flow Diagram of Fluid Cracking Unit 757
21.2 An Analysis of Particle Size Distribution in a Gas Stream to Electrostatic Precipitator for One Type of Catalyst 759

Figure No. Page No.

21.3 Electrical Resistivity of One Particular Type Precipitator Inlet Catalyst Dust (23% Moisture Content by Volume) - Measured in Laboratory.... 760
21.4 A Single-Stage Vertical Wire and Pipe Unit 762
21.5 Oil Shale Retorting Process 764
21.6 Acetylene (Wulff Process)...................... 765
21.7 A Process Flow Diagram of Production of Carburetted Water Gas 767
21.8 Flow Diagram for Typical Coal Carbonization Process....................................... 769
21.9 Installed Precipitator Gas Volume Trend in Catalytic Cracking Units Over 1940-1967 772
21.10 Installed Precipitator Gas Volume Trend in Carburetted Water Gas Over 1940-1958.......... 774
21.11 Cost of Detarring Precipitators with 95% Collecting Efficiency (1959-1969) 778

22.1 Nonferrous Metals Production in the United States 789
22.2 Use of an Electrostatic Precipitator on Ore Sintering Machine Exhaust Gas 790
22.3 Purification System for Soderberg Cell Gases 800
22.4 Flow Diagram for Aluminum Production......... 801
22.5 Flow Diagram Showing Steps in the Copper Extraction and Refining Process 811
22.6 Schematic Diagram Showing Electrostatic Precipitators Used for Blast Furnace Gas 816
22.7 Electrostatic Precipitator Operating on a Rotary Furnace............................ 817
22.8 Flow Diagram for a Zinc Plant Utilizing a Retort Process for Zinc Reduction 820
22.9 Collection Efficiency as a Function of Specific Collection Electrode Area (ft^2/cfm) for Nonferrous Installations 824

23.1 Schematic Diagram of Pilot Plant for Incinerator Gas Turbine System Utilizing Electrostatic Precipitator for Fine Cleaning of Turbine Gas.... 839

Figure No. Page No.

24.1 Performance of Typical Mechanical Cyclone Dust Collector.................................. 846
24.2 Purchase Cost of Dry Centrifugal Collector...... 849
24.3 Purchase Cost of Wet Collectors 850
24.4 Purchase Cost of High-Voltage Electrostatic Precipitators.................................. 851
24.5 Purchase Cost of Fabric Filters................ 852
24.6 Installed Cost of High-Voltage Electrostatic Precipitators.................................. 853
24.7 Comparative Costs for Particulate Control for 100,000 cfm Installation........................ 855
24.8 Total Shipments of Various Types of Particulate Control Devices for 1963 and 1967.............. 857

LIST OF TABLES
PART II

Table No. Page No.

15.1 Energy Sources for Electric Power Generating Plants 349
15.2 Particulate Emission from Heat and Power Generation before Collection (Millions of Tons per Year)........ 350
15.3 Typical Flue Gas Compositions of Selected Components Measured at Moisture Electrostatic Precipitator Inlet 364
15.4 Variations in Precipitation Rate Parameter Predicted by the Regression Analysis Equation for Fly Ash Precipitators 394
15.5 Fly Ash Precipitator Installations 395
15.6 Precipitator Costs (1960 through 1964 and 1965 through 1969) 402
15.7 Average Fly Ash Precipitator Costs from 1923 through 1969 406
15.8 Summary of Operating Costs for Fly Ash Precipitators 409
15.9 Operating Costs for Precipitators for Electric Utilities, Gas Flow Rate - 200, 000 acfm 410
15.10 Operating Costs for Precipitators for Electric Utilities, Gas Flow Rate - 500, 000 acfm 411
15.11 Operating Costs for Precipitators for Electric Utilities, Gas Flow Rate - 1 million acfm 412
15.12 Operating Costs for Precipitators for Electric Utilities, Gas Flow Rate - 2 million acfm 413

16.1 Summary of Basic Pulping Processes 419
16.2 A Survey of Chemical and Semichemical Mills in the United States 420
16.3 Summary of Chemical Cooking Processes 423
16.4 Summary of Pulp Mill Recovery Systems 447
16.5 Electrostatic Precipitator Costs (1960 through 1969) . 453
16.6 Precipitator Costs (1965 through 1969) 471

17.1 Typical Coke Oven Gas Precipitator Data 488
17.2 Typical Coke Oven Gas Precipitator Design Parameter 489
17.3 Range of Composition of Particulate from Sinter Plants 492
17.4 Use Statistics on Sintering Machines, 1951-1968 497

Table No. Page No.

17.5 Performance Parameters for Precipitators Collecting Sinter Machine Dust 499
17.6 Sinter Plant - Windbox - Electrostatic Precipitator .. 504
17.7 Chemical Analysis of Blast Furnace Flue Dusts 509
17.8 Weight Percent Composition of Dust Samples from Blast Furnace Flue Dust 510
17.9 Size Analysis of Blast Furnace Flue Dust 511
17.10 Distribution of Total and Electrostatically Controlled Gas Volume of Effluent from the Blast Furnace from 1931 to Present 515
17.11 Fume Emissions from Open Hearth Furnaces (Adapted from Reference 4) 531
17.12 Variation of Fume Generation with the State of the Heat in Open Hearth Furnaces (Adapted from Reference 4) 533
17.13 Open Hearth Dust Analyses (from published data) 534
17.14 Analysis of Fume from Oxygen-Blown Steel Manufacturing Processes 535
17.15 Size Distribution of Fumes in an Open Hearth Effluent (Adapted from Reference 23) 537
17.16 Waste Gases from Open Hearth Furnace Fuels (Adapted from Reference 23) 540
17.17 Precipitator Installations on Open Hearth Furnaces 1949-1968 544
17.18 Open Hearth Furnace - Electrostatic Precipitator ... 549
17.19 Particle Size Analysis of BOF Fume from Two Installations 554
17.20 Composition of Fume from BOF Data Shown in Table 17.19 554
17.21 Precipitator Installations on Basic Oxygen Furnaces 1956-1968 560
17.22 Application and Design Variables for Two BOF Furnace Precipitators 563
17.23 Basic Oxygen Furnace - Electrostatic Precipitator (Reference 14) 564
17.24 Basic Oxygen Furnaces - Electrostatic Precipitators (Reference 14) 566
17.25 Electric Arc Steel Furnace Fume Emission Data 571
17.26 Chemical Analysis of Fume Samples 573

Table No.		Page No.
17.27	Chemical Analysis Range of Electric Furnace Dust	575
17.28	Qualitative Spectrochemical Analysis and Concentration Estimates of Detected Constituents	576
17.29	Electric Furnace Dust Composition (Reference 44)	577
17.30	Electric Steel Furnace Fume Particle Size Data	579
17.31	Collection of Electric Arc Steel Furnace Fumes (Reference 20)	583
17.32	Electric Arc Furnace - Electrostatic Precipitator	585
17.33	Precipitator Installations on Scarfing Machines (Period 1954-1969)	587
17.34	Design and Performance Data for Electrostatic Precipitators Collecting Dust from Scarfing Machine	589
17.35	Electrostatic Precipitators for Scarfing Machines (Reference 14)	590
17.36	Electrostatic Precipitators for Scarfing Machines (Reference 14)	591
17.37	Chemical Composition of Cupola Dust	596
17.38	Dust Content of Cupola Waste Gases	596
18.1	Precipitator Inlet Dust Loadings for Cement Industry Precipitators	619
18.2	Range of Gas Moisture Contents for Precipitators in the Cement Industry	622
18.3	Precipitator Inlet Gas Temperatures	624
18.4	Electrostatic Precipitator Installations in Cement Plants Period 1912 through 1969	626
18.5	Summary of Precipitator Design Efficiencies (1949-1969)	629
18.6	Summary of Precipitator Design Gas Velocities for Cement Industry Precipitators (1949-1969)	630
18.7	Summary of Precipitator Design Input Powers for Cement Industry Precipitators (1949-1969)	631
18.8	Cement Industry Summary of Statistics (1949-1969) Precipitator Design Avg. Field Strength	632
18.9	Summary of Performance Statistics on Cement Plant Electrostatic Precipitators	633
18.10	Design Data for Two Current Installations	636
18.11	Design and Performance Precipitation Rate Parameters for Cement Plant Electrostatic Precipitators	637

Table No.		Page No.
18.12	Maintenance and Operating Costs for Precipitators Installed in Cement Plants	642
18.13	Precipitator Inlet Dust Loadings for Gypsum Industry Precipitators (1935-1969)	645
18.14	Range of Gas Moisture Contents to Precipitators in Gypsum Plants (1935-1969)	647
18.15	Summary of Precipitator Gas Temperatures for Gypsum Industry Precipitators (1935-1969)	650
18.16	Electrostatic Precipitator Installations in Gypsum Plants (Period 1935-1969)	651
18.17	Electrostatic Precipitator Installations in Gypsum Plants Listed by Area of Application (Period 1935-1969)	652
18.18	Summary of Performance Data on Gypsum Plant Electrostatic Precipitators	654
18.19	Gypsum Industry Economics	657
18.20	Maintenance Data for Precipitators Installed on Gypsum Calciners	658
19.1	Sulfuric Acid Mist Precipitator Installations 1945-1969	676
19.2	Summary of Sulfuric Acid Mist Precipitator Costs 1960-1969	681
19.3	Average Cost per acfm for Sulfuric Acid Mist Precipitators	683
19.4	Comparison Between Design and Performance Parameters for an ESP Operating on Acid Mist from a Sulfuric Acid Plant	685
19.5	Hot Phosphorus Precipitator Installations 1938-1969	692
19.6	Design Efficiency Trends for Hot Phosphorus Precipitators (1938-1969)	696
19.7	Summary of Design Data Electric Furnace Phosphorus Precipitator (1938-1969)	697
19.8	Summary of Design Data on Two Precipitators for Cleaning Phosphorus Furnace Gas	701
19.9	Electric Furnace Phosphorus Precipitator Costs	702
19.10	Average Electric Furnace Phosphorus Precipitator Costs (1938-1969)	703
19.11	Phosphoric Acid Mist Precipitator Installations 1927-1956	708

Table No.		Page No.
19.12	Phosphoric Acid Mist Precipitator Installations 1927-1956	709
19.13	Precipitator Gas Velocity (Design) Phosphoric Acid Mist Precipitator 1927-1969	713
19.14	Precipitator Inlet Temperature (Design) Phosphoric Acid Mist Precipitator 1927-1969	714
19.15	Precipitator Inlet Loading (gr/scf) (Performance) Phosphoric Acid Mist Precipitator 1927-1969	716
19.16	Range of Performance Data Phosphoric Acid Mist Precipitator	720
19.17	Phosphoric Acid Mist Precipitator Cost	721
19.18	Carbon Black Precipitator Installations 1942-1956	725
19.19	Carbon Black Precipitator Cost Data (1942-1952)	726
19.20	Range of Performance Statistics Carbon Black Precipitators (1942-1956)	727
20.1	Range of Plant and Furnace Capacities for Municipal Incinerators Installed and Rebuilt from 1945-1965 - U. S. and Canada	730
20.2	Refuse Composition of Selected Countries (weight percentage)	731
20.3	Incinerator Heating Value	733
20.4	Incinerator Emission Properties - United States	738
20.5	Analysis of Gases from Incinerator Grates	739
20.6	Electrostatic Precipitators Installed on Municipal Incinerators in North America	741
20.7	Municipal Incinerator Installation	748
20.8	Summary of Performance Data	749
20.9	Estimated Operating Costs for Electrostatic Precipitators for Municipal Incinerator Dust Collection	752
21.1	A Summary of Application Data on Electrostatic Precipitator Installations in Catalytic Cracking Units 1940-1967	771
21.2	A Summary of Application Data on Electrostatic Precipitators on Various Applications for Removing Tar and Oil Mist 1940-1963	773
21.3	FOB Cost of Precipitators for Catalytic Cracking Units 1951-1962	775

Table No.		Page No.
21.4	Cost of Detarring Precipitator in Fuel Gas Area 1945-1956	776
21.5	Cost of Detarring Precipitators in Fuel Gas 1959-1969	777
21.6	A Summary of Application Data for Precipitators in Fluid Catalytic Cracking Application 1951-1962	780
21.7	A Summary of Performance Data in Fuel Gas Application 1940-1956	781
21.8	Field Tests on Fluid Catalytic Cracking Units	782
21.9	Design and Performance Data for Two Powdered Catalyst Precipitators	783
22.1	Design Performance Data for Aluminum Reduction Furnace Precipitators	802
22.2	Operating Data of Multiple Hearth Roasting Furnaces	804
22.3	Fluid Bed Roasters	805
22.4	Operating Data for Reverberatory Furnaces	807
22.5	Operating Data for a Copper-Matte Converter	809
22.6	Operating Data for Sintering Installations	813
22.7	Operating Data for Blast Furnaces	814
22.8	Typical Zinc Roasting Operations	819
22.9	Typical Zinc Sintering Operations	821
22.10	Summary of Electrostatic Precipitators in Western Smelters Prior to 1920	823
22.11	Precipitator Performance in the Nonferrous Metallurgical Industries	825
22.12	Electrostatic Precipitator on Roasters Period 1923 through 1969	826
22.13	Summary of Electrostatic Precipitators for Cadmium Recovery Zinc Sinter Machines (1935 through 1955)	827
22.14	Summary of Electrostatic Precipitators for Miscellaneous Nonferrous Applications 1924 through 1968	828
22.15	Summary of Design and Application Parameters on Electrostatic Precipitators for the Aluminum Industry (1940 through 1967)	829
22.16	Cost Data for Electrostatic Precipitators for Cleaning Ore Roaster Gas	830

Table No.		Page No.
22.17	Economics on Electrostatic Precipitators for Cadmium Recovery Zinc Sinter Machines	831
22.18	Economics on Electrostatic Precipitators in the Aluminum Industry	832
22.19	Performance Statistics on Roaster Gas Electrostatic Precipitators (1933-1953)	834
22.20	Performance Statistics on Cadmium Recovery Zinc Sinter Machine Gas Precipitators (1935-1955)	835
23.1	Projections from Pilot Precipitator Tests for Cleaning of Turbine Gas	841
23.2	Gas Pipeline Precipitator Design and Operating Data	842
24.1	Comparison between Particulate Control Costs for a 100,000 cfm Unit (Reference 3)	854
24.2	A Summary of Electrostatic Precipitator Applications	859
24.3	Sales of Dust Control Equipment Used in Various Processes during the Period 1966-1967	860

PART II - APPLICATION AREAS

DESCRIPTIVE SUMMARY OF PART II

Part II of this manual is a review of the application of electrostatic precipitators in each of eight major application areas. In addition, an analysis is presented of the use of dust control equipment in a number of manufacturing operations and the factors influencing the use of electrostatic precipitators in new dust control applications.

Within each application area, a review of the process is given with particular emphasis on the dust and gaseous emissions. This is followed by a tabulation of input and design parameters for precipitators operating on various types of dust control problems and an analysis of critical design parameters and test results. Cost data are also presented to show the range of FOB and erected precipitator costs for various efficiency levels and gas volumes.

The information presented has been derived from a review of the literature pertaining to the particular application, a tabulation of information by Research-Cottrell, Inc., of design, cost, and input parameter data, and, finally, an analysis of design and test data obtained from a survey of operating installations. The following is a summary of the information presented in greater detail in the following chapters of the manual.

ELECTRIC POWER GENERATION

Electrostatic precipitators are used in the electric power generation industry principally for the control of emissions from coal-fired steam generating plants. Although there is a trend toward the use of nuclear power generation, the expected increase in the total power consumption and the upgrading of existing control equipment is expected to result in the expanded use of electrostatic precipitators in this application for a number of years.

Fly ash is generated from both pulverized-coal-fired boilers and cyclone boilers. The character and amount of fly ash from these two types of boilers vary with the type and chemistry of the coal being burned and the particular operating conditions of the boiler. The ash content of the coals being burned varies from 5-25%, and, together with the ash-fusion temperature and boiler operation, determines the dust load that must be handled. Typical dust loads range from 2-7 gr/cu ft of gas.

The particle size distribution of fly ash varies with the type of boiler and type of coal. For a pulverized fuel boiler, the mass median diameter of the fly ash is around 10-15 microns. For a cyclone furnace, the mass median diameter varies over a wider range and is generally smaller (5-15 microns). It is rather common practice to precede fly ash precipitators with mechanical collectors which remove mainly the large size fraction of the dust. The mass median diameter of the fly ash to a precipitator preceded by a mechanical collector is around 3 microns.

The resistivity of the collected fly ash is perhaps the most important property influencing collection by electrostatic precipitators. If the resistivity is high (above about $2x10^{10}$ ohm-cm), the voltage and current to the precipitator must be kept low to prevent excessive sparking and back corona. Under these conditions, the charge acquired by the dust will be low, the charging time will be long, and the collection field will be low. Consequently, the performance of the precipitator will be impaired.

At the other extreme, too low a resistivity will permit reentrainment of the collected dust and result in low efficiency. There is an optimum resistivity therefore for maximum precipitator performance.

Resistivity of fly ash is determined by the temperature of the flue gas and the chemistry of the coal. At temperatures of around 450-500° F or above, volume conductivity predominates and the resistivity is always below the critical $2x10^{10}$. As the temperature decreases, resistivity increases. This trend would continue under bone-dry conditions. However, at temperatures in the range of 300° F, moisture in the flue gas is adsorbed on the fly ash surface and alters its resistivity by a mechanism called surface conduction. The lower the temperature, the greater the rate of adsorption, so that resistivity continues to decrease with decreasing temperatures.

In addition to temperature, the percentage of sulfur in the coal also influences fly ash resistivity. Studies of fly ash resistivity indicate that the SO_3 present in the flue gas acts to alter the rate of moisture adsorption and serves as a secondary conditioning agent. On the average, the amount of SO_3 present in the flue gas is directly related to the sulfur content of the coal. However, operating conditions in the boiler, and perhaps other constituents of the fly ash, govern the quantity of sulfur appearing as SO_3.

In addition to the temperature and the amount of SO_3 present in the gas, the chemistry of the fly ash appears to influence the adsorption rate.

A basic ash is reported to contribute to a higher rate of adsorption of SO_3. Thus, even though there is a good statistical correlation between sulfur content of the coal and fly ash resistivity, the variations for a single condition are too great to permit accurate prediction of resistivity based on sulfur content alone.

Measurement of resistivity of fly ash should be made in situ, utilizing the techniques described in Chapter 12 of Part I.

Fly ash precipitators are generally designed on the basis of the Deutsch-Anderson equation relating efficiency to gas volume and collecting surface area. Experience with large numbers of precipitators has shown, however, that the precipitation rate parameters can vary between around 3 cm/sec to 17 cm/sec. The major problem in the design is therefore to narrow the range of uncertainty in selection of the precipitation rate parameter.

The value of the precipitation rate parameter for most fly ash precipitators is around 10 cm/sec (0.33 ft/sec). Variations from this value can occur if the fly ash properties are either more favorable or more adverse than normal.

Low values of precipitation rate parameter are normally associated with excessive gas velocities, uneven gas flow distribution, high dust resistivity, low dust resistivity or improper rapping. Problems of gas flow distribution and gas velocity can be handled by proper design through the use of models, etc. These problems are perhaps made more severe in the case of fly ash precipitators by the large gas volumes that must be handled.

High resistivity problems are usually associated with high ash, low sulfur coals. Low resistivity problems can occur if the gas temperature is low and high sulfur coal is being burned.

Problems with high fly ash resistivity can be resolved by several approaches which include: (1) increasing precipitator size, (2) changing the flue gas temperature, or (3) adding chemical conditioning agents.

The first of these alternatives is straightforward; however, about three times the normal collection surface area may be required and the costs of this alternative may be prohibitive.

Precipitators are normally located downstream of the air heater and operate at temperatures in the range of 250-350° F. However, they can be located ahead of the air heater where gas temperatures are in the vicinity of 700° F. At these temperatures, the dust resistivity is determined principally by volume conductivity and is independent of the sulfur content of the coal. Since gas volumes are about 1.5 times those at 300°F, additional precipitator capacity is required. However, this alternative may be attractive for many applications.

Operation of precipitators following the air heater can utilize the option of lowering the flue gas temperature to reduce dust resistivity. Reduction of temperatures from 300° F to 260-270° F can reduce fly ash resistivities by factors of 10 or more, and can minimize the high resistivity problem. Because resistivities can change so rapidly within this temperature range, control may become a problem.

Use of chemical additives to control dust resistivity is a third alternative. Small additions of SO_3 (up to 10-15 ppm) to the flue gas have altered dust resistivity by factors of 10 or more and resulted in substantial improvements on precipitators which are limited by sparking due to high dust resistivity. Other chemical additives, such as H_2SO_4, are being evaluated as conditioning agents.

Problems with low dust resistivity have been encountered when burning high sulfur coal and operating at low flue gas temperatures. The difficulty has been identified as excessive scouring or reentrainment due to the reduction in force holding the dust to the collection plate. The problem is associated with gas velocity and is apparent when relationships between gas velocity and precipitation rate parameter are plotted for various resistivities. This problem can be resolved by increasing the flue gas temperature, injection of ammonia, reduction in gas velocity, or a combination of these effects.

In addition to determining the collection surface area, design of a fly ash precipitator also includes determining the power supply requirements. It has been shown empirically that efficiency is related to the corona power delivered to the precipitator. Curves showing this relationship for fly ash precipitators show a good correlation and are useful in both design and troubleshooting of fly ash precipitators.

The number of independently energized bus sections is another design variable. Since fly ash precipitators are generally designed to handle large gas volumes, the degree of sectionalization is of greater importance than in some other applications. If very high collection efficiencies are desired, a higher degree of sectionalization should be used. The advantages are: (1) higher operating voltages in the spark-limited mode, (2) lower internal impedance with better spark quenching, and (3) less percentage of the precipitation would be disabled by the outage of a single section.

Since a number of factors influence precipitator performance, and since these vary between installations, a technique of regression analysis has been developed by several investigators exploring precipitator operation. A method of relating the expected precipitation rate parameter with the significant parameters has been developed to serve as a guide to analysis of precipitator specifications. The technique gives a correlation coefficient of about 0.85. A second regression analysis performed on data from a group of installations gave a correlation coefficient of around 0.9 when using more fundamental precipitator and dust parameters.

PULP AND PAPER

Electrostatic precipitators are used in the pulp and paper industry to remove particulates carried by the effluent gases from black liquor recovery boilers.

The kraft or sulfate process constitutes approximately 55% of the total production in this country. Of these kraft mills, about 65% have electrostatic precipitators installed on the recovery boiler.

The economics of the kraft process requires the recovery and reuse of the spent chemicals used in the cooking process and recovery of the heat content of the concentrated spent liquor. This is accomplished by combustion in the recovery boiler which releases large volumes of flue gases. During combustion, a significant fraction of the recoverable chemicals is entrained as particulates with the flue gas. Recovery of these particulates constitutes a significant economic factor as well as a means for controlling air pollution.

Particulate emissions from the recovery boiler are extremely fine hydroscopic particles, composed principally of sodium sulfate and sodium carbonate with small quantities of sodium chloride, sulfide, and sodium sulfite. Because of its hydroscopic nature, sampling to determine particle size distribution is difficult. A technique for precipitating a sample onto a copper mesh electron microscope target and subsequently counting the particles in various size ranges has been developed. The collected sample must be encapsulated in a protective atmosphere to prevent moisture pickup during transfer from the duct to the microscope. The mass median particle size for recovery furnace dust is around 1.9 microns. By count, the median size is about 0.4 micron.

Electron photomicrographs of the dust samples from recovery boilers show a change in the character of the dust dependeing upon operating temperature. At temperatures in the range of 350-360° F, the dust is primarily spherical particles. At a temperature of about 280-290°F, the dust contains large numbers of needle-like particles. These differences in particulate structure are thought to be the cause of the variations in the transporting and rapping properties of the collected dust.

Resistivity data for recovery furnace dust have not been reported extensively, primarily because high resistivity has not been identified as a problem.

Electrostatic precipitators for recovery boilers are of a variety of types including vertical- and horizontal-flow types, wet and dry bottom types, and steel, tile or concrete shell types.

Dry bottom precipitators have been used extensively in recovery boiler precipitators from 1930 to 1948. In the dry bottom precipitators, dust from the collection plates was collected in pyramidal-type hoppers. The wet bottom precipitator was introduced to minimize problems associated with the collection, transporting, and redissolving of the collected salt cake. In the wet bottom precipitator, a solution of 50% liquor is pumped into the collection hopper. The collected dust is rapped from the plates and falls into the liquor solution, where it is dissolved and subsequently removed.

Current emphasis on odor control from kraft mills has resulted in a trend toward elimination of the contact between the flue gas and black liquor. Consequently, wet bottom precipitators may not be used on new kraft process mills. Also, elimination of the direct contact evaporators may influence the properties of both the gas and particulates to be handled by electrostatic precipitators.

Precipitators for recovery furnace boilers are designed for dust concentrations in the range of 1-9 gr/acf, with the bulk of the installations in the range of 2-5 gr/acf. Inlet temperatures range from 225-375°F, with the majority of installations operating at temperatures of 275-325°F. Gas velocities range from 2-8 ft/sec, with the largest number of precipitators operating in the 3-6 ft/sec range.

Connected input power for the majority of installations is in the range of 100-200 watts/1000 cfm. Design field strengths range from around 7-16 kV/in., computed as the ratio of average design voltage to plate spacing.

Design precipitation rate parameters for recovery boiler precipitators range from around 0.2 to 0.35 ft/sec.

IRON AND STEEL INDUSTRY

The application of electrostatic precipitators in the iron and steel industry has been in the cleaning of gaseous effluents from steelmaking furnaces, blast furnaces, foundry cupolas, sinter machines, and by-product coke ovens.

Steelmaking processes have undergone a number of changes in production methods. One of the earliest steelmaking furnaces was the Bessemer converter. About 1868, the open hearth process was invented and was the primary method of steelmaking for a number of years. The basic oxygen furnace, introduced around 1950, has increased in importance, and in 1969 about 40% of the total steel made was produced in BOF furnaces.

By-product coke gas. Production of metallurgical coke is made by the beehive process and in by-product coke ovens. The latter accounts for about 98% of all the metallurgical coke produced in this country.

Precipitators are used in the by-product coking process to remove the tars and particulate matter from the gases prior to recovery of the hydrocarbons.

The precipitator most often used for detarring is a cylindrical type consisting of collection electrodes made from 6-8 in. pipe, 6-9 ft long suspended from a header plate in a round shell. Discharge electrodes are suspended axially through the cylinders. The precipitators are maintained at a temperature high enough so that the collected tar drains from the plates and no rapping is required.

Three problems associated with the use of precipitators for cleaning coke oven gases are as follows: (1) Collection of the fluid tar makes electrical insulation difficult and requires that insulators be kept out of the gas stream. The insulators are also heated and cleaned on a regular scheduled basis. (2) The gases being cleaned are combustible when mixed with air, so that no inleakage of air must occur. Design often includes operation at positive pressure to insure no inleakage. (3) The gases can be corrosive. Protection against corrosion is provided by the film of tar on the interior surface of the collection electrodes. Spraying of tar on the exterior collection electrode surface can minimize corrosion at that point.

Coke oven gas precipitators are designed to handle gas volumes in the range of 5-15, 000 acfm with efficiencies in the range of 95-98%. Precipitation rate parameters are about 0. 2 - 0. 3 ft/sec. Inlet loadings are in the vicinity of 0. 5 gr/acf. Gas velocities average around 8 ft/sec, which is higher than for some applications since reentrainment is not a problem. Power densities are 8-10 watts/ft^2.

Sinter plants. Sintering is a process for agglomerating iron-bearing fines to prevent their loss during reduction in the blast furnace.

The raw material is composed of iron-bearing fines, coke or coal dust, and a fluxing material such as limestone or dolomite. As the material moves through the sinter stand, it is ignited by surface burners and combustion is sustained by air drawn through the mixture by fans. The products of combustion are collected in a group of compartments called windboxes located beneath the grates of the sinter machine.

The particulate material emitted from the sintering process is a result of the mechanical handling of the raw material and combustion of the coal or coke. Under normal conditions, 5-100 lb of dust is produced per ton of sinter produced. Gas volumes vary from 100, 000-450, 000 cfm, with dust loadings of 0. 5 - 6. 5 gr/scf. About 80-90% of the particulate is greater than 20 microns. The dust contains fewer fines than metallurgical furnace operations. However, when preceded by a mechanical collector, the dust can be relatively small. Typically, the mass median diameter of the particles following a mechanical collector would be around 6 microns, whereas without a mechanical collector, the mass median diameter would be around 50 microns.

The electrical resistivity of sinter machine dust can vary radically depending upon the type and amount of fluxing material in the burden makeup. The amounts of fluxing material can vary from around 10-35%, and this variation can change the resistivity of the dust by several orders of magnitude. The very high resistivities associated with the higher percentages of fluxing materials can result in excessive sparking, lower operating voltage and power, and generally poorer precipitator performance.

The effect of the addition of the flux has not been fully explored; however, the condition is similar to that encountered in fly ash precipitators where limestone is added to the boiler for removal of sulfur oxides. It is postulated that the limestone reacts preferentially with the SO_3 present in the gases. Since the SO_3 is a secondary conditioning agent, surface conductivity would be decreased.

Studies of the mechanisms of conduction and possible corrective measures have not been explored for sinter machine precipitators to the same extent that they have for fly ash precipitators.

Design of sinter machine precipitators has conventionally been based on conditions in which resistivity has not been a problem. Because of the relatively large particle size of the dust, the range of design precipitation rate parameters has been from around 0.25 to 0.4 ft/sec. However, test precipitation rate parameters as low as 0.08 ft/sec have been obtained when high dust resistivity has been encountered.

Other design parameters for sinter machine precipitators are: gas velocity - 4-5 ft/sec, temperature - 250-300° F, electric field - 8-10 kV/in., average inlet dust loading - 1 gr/acf, and average precipitator power - 70 watts/1000 cfm.

Blast furnace. The effluent gases from blast furnaces have heating values of about 100 Btu/scf, which makes them valuable as a fuel for heating. The gas, however, contains particulate material carried over from the blast furnace and must be cleaned to prevent clogging of gas burners and gas mains. Electrostatic precipitators have typically been used for this service.

Blast furnace gas is usually passed through a dust catcher where the heavier particles are separated by inertial forces. From the dust catcher, the gas is passed to a wet scrubber and an electrostatic precipitator.

Precipitators for cleaning blast furnace gas are conventionally of the vertical-flow type employing cylindrical collection electrodes, although horizontal-flow, plate-type precipitators are also used.

Since the gas delivered to a blast furnace precipitator is cooled to saturation by the preceding wet scrubber, a wet type precipitator is used. Particulate removal is through a slurry hopper. The slurry is normally piped from the hopper to a dust disposal system.

The use of precipitators for cleaning blast furnace gases has generally followed the blast furnace production capacity. However, in recent years, the sales of blast furnace gas precipitators have been low compared with the peak periods during the 1940-1960 period.

Inlet gas temperature for blast furnace gas precipitators is low as compared with other applications. Typical temperatures are in the range of 70-100° F. Inlet dust loadings are also low, of the order of 0.05 - 0.3 grains/scfd for most installations.

Gas velocities are relatively high, around 6-15 ft/sec, since reentrainment is not a serious problem with wet precipitators. The average field strength (average voltage-to-electrode-spacing ratio) varies from around 9-15 kV/in. Input power ranges from 50-300 watts/1000 cfm.

Design precipitation rate parameters for blast furnace gas precipitators are around 0.2 - 0.4 ft/sec. Test data vary over a considerably wider range.

Open hearth. Open hearth steelmaking furnaces consist of a large refractory lined disk into which metal from the blast furnace, steel scrap, iron ore, and limestone are charged. Heat is provided by furnace burners which burn oil, natural gas or tar with combustion air that is heated in regenerative heat exchangers called checkers. Open hearth furnaces are often equipped with oxygen lances to facilitate oxidation of the carbon and other elements to be removed.

The particulate material carried out by the exhaust gases comes from a variety of sources, including dirt and other fines on the charge material, oil and grease, and volatile metal oxides from the scrap charge. During oxygen lancing, large amounts of iron oxide are evolved together with lesser amounts of nonmetallic oxides from the slag. The quantity of exhaust gases, particulate loading, and particulate composition vary widely during the period of the heat. Consequently, the precipitator input conditions vary depending upon operating conditions of the furnace.

The sizes of the particulate matter from open hearth furnaces range from less than 0.03 micron to several microns. Composite samples representing dust evolved during the entire heat indicate that 50% of the particulate is less than 5 microns. However, during the lime boil, the dust is considerably finer with as much as 77% less than 5 microns and 20% less than 1 micron. Resistivity of open hearth furnace dust varies with moisture, temperature, and composition.

Precipitators for cleaning open hearth gases are generally the horizontal-flow, duct-type with a steel shell. In large steelmaking shops, a multiplicity of furnaces are serviced by a common air pollution control system, which may consist of mechanical collectors, washers, and electrostatic precipitators singly or in combination. Open hearth precipitators are designed with precipitator rate parameters in the range of 0.15 - 0.3 ft/sec.

The total installed capacity of open hearth precipitators has steadily increased since 1950. In 1969, the total capacity reached approximately 12 million acfm.

Basic oxygen furnaces. The basic oxygen process of steelmaking utilizes high pressure oxygen introduced into the mouth of a basic, refractory lined converter to oxidize carbon and other elements from the charge. There is no external source of heat in the converter, and molten pig iron from the blast furnace constitutes the major constituent of the furnace charge. After charging, oxygen is blown into the converter at high pressure through a water-cooled lance.

Gas and dust emissions from BOF converters vary greatly with the stage of heat. During charging, the gaseous and particulate emission levels are low. During the oxygen blow, which may last for 20 minutes, large amounts of fume and gas are evolved. Gas volumes range from 200,000 - 1,200,000 cfm at temperatures of 3000 - 3200°F and may carry 300 lb/min or more of dust.

Fume from the basic oxygen furnace is composed primarily of iron oxide in amounts of 20-60 lb/ton of steel. Fume concentrations of up to 15 gr/scfm are produced under peak conditions. The fume is finely divided with most of the particles ranging from 0.1 - 1 micron in size, although on a weight basis, a considerable portion of the burden exists as relatively large particles.

Resistivity of BOF dust is a function of moisture content and temperature. Moisture comes from the evaporation of water in the cooling tower located between the precipitator and the furnace.

The cooling tower is an important part of the gas cleaning facility. It should be large enough to insure complete evaporation of the water to prevent carryover and subsequent clogging of the precipitator. Since the gas temperature varies, the rate of heat transfer varies, so that water quantities to the cooling tower must be controlled.

Precipitators for BOF gas cleaning are generally the horizontal-flow, duct-type with steel shells. Design precipitation rate parameters vary from around 0.15 - 0.25 ft/sec. The generally smaller particle size distribution accounts for the lower values of precipitation rate parameter.

Gas temperatures for BOF precipitators range from around 250-550° F depending upon design philosophy. Design efficiencies are generally 99+% for newer installations.

One problem associated with BOF precipitators is that the collection efficiency at the start of the oxygen blow is reduced. Cooling towers are generally set to operate the water sprays when the gas temperature reaches around 500° F. In the interim, dust resistivity is high and emissions from the stack can be relatively high during this period. The condition is called a lance puff and can be controlled by introduction of steam to condition the dust during the period before the cooling tower sprays come on.

Electric arc furnace. The electric arc furnace as used in the steelmaking process consists of a refractory lined structure with a dish-shaped bottom and a domed roof. Steel scrap, and perhaps hot metal from a blast furnace, are charged into the furnace and heated by an electric arc developed between graphite electrodes which are lowered into the furnace charge.

After meltdown, oxygen is introduced to remove the carbon and other elements. Oxygen sources can be from various sources including oxygen gas, oxides of alloying elements, iron ore, decomposition of limestone, etc.

The flue gases from electric furnaces contain large concentrations of carbon monoxide which must be converted to CO_2 prior to entering the precipitator to minimize the explosion hazard. This is usually accomplished by admitting air to the hot gas stream and allowing combustion to take place in the high temperature region. After combustion, the gases are passed through a cooling tower to reduce the temperature before entering the gas cleaning equipment.

The particulate emissions from electric arc furnaces average around 5-30 lb/ton of steel produced. The size and composition of the dust emitted from electric furnaces vary with the type and cleanliness of scrap and the metal refining procedure. Nonferrous impurities in the scrap can lead to significant quantities of oxide fumes. Also, presence of oil and grease in the scrap can produce large amounts of carbonaceous particulate matter during early stages of meltdown.

Dust from electric arc furnaces tends to be extremely fine. Data on particle size distribution vary. Some sources indicated that as much as 90-95% of the fume is below 0.5 micron. Other data would indicate a somewhat coarser dust.

Electrostatic precipitators have been used to clean electric furnace gases in the United States and Europe. However, the small size of the particles necessitates the use of a large precipitator to achieve high collection efficiencies, and other methods of cleaning electric furnace gases have been used to advantage in some instances.

Design precipitation rate parameters for electric arc furnace precipitators range from around 0.12 - 0.16 ft/sec.

Scarfing machines. Scarfing is the operation of removing the skin of the slab in the production of steel. In the scarfing machine, the slab passes under cutting torches which burn through the slab. The process generates iron oxide particles and fume which constitute an emissions problem. The application of electrostatic precipitators to this service is limited; the total capacity installed being around 500, 000 cfm.

Cupolas. The iron cupola is used to provide a source of molten metal for cast iron foundries. The cupola is a refractory lined, cylindrical furnace which, when charged with pig iron, scrap, coke, and flue, provides a self-sustaining exothermic reaction to melt the charge and maintain it at the proper temperature.

Particulate emissions from iron cupolas consist primarily of oxides of iron, silicon, calcium, aluminum, magnesium, and manganese. The dust load in the effluent gases varies widely with ranges from 0.9 - 6.5 and 1.3 - 11 grains/scfm reported for cold and hot blast cupolas, respectively.

Particle size distribution of cupola dust varies between wide limits depending upon melt rate, type and cleanliness of scrap, and furnace operating conditions. Values range between 10% of the dust less than 10 microns to 40% less than 1 micron.

The gases from the cupola are high in carbon monoxide and must be burned prior to entering an electrostatic precipitator because of the potential explosion problem. Burning normally is accomplished by introduction of combustion air at the exit of the cupola. The combustion process can also burn carbonaceous particulates, and thus can have an influence on the particle size distribution as well as the composition.

Use of electrostatic precipitators on iron cupolas in this country has been extremely limited. Poor experience with operation of precipitators on a few installations has influenced the choice of dust collection equipment to a considerable extent. Also, the extreme variability of cupola operation resulting in extremely wide ranges of particle size and composition has tended to limit use of precipitators for this application.

ROCK PRODUCTS

Electrostatic precipitators are used in the rock products industry in the collection of dust from cement kilns, gypsum kettles, and from auxiliary grinding, transporting, and handling operations.

Portland cement. Portland cement is produced by a wet or dry process which defines the conditions under which the ingredients are ground and fed into the calcining kiln. There are variations of these two production methods which are often termed semi-wet or semi-dry processes.

In the basic cement production process, the raw materials, consisting of lime, alumina, iron oxide, and a fluxing material are ground in a mill and introduced in a kiln. The material fed into the kiln is dried in the initial section, calcined as it passes down the kiln, and finally fused into a clinker in the final section of the kiln. The clinker is then removed, cooled, and ground to produce the final product.

The effluent gases from the kiln range from 40,000-100,000 cfm depending upon the type of kiln, method of gas cooling, and the method of preheating the raw materials. The gases are composed of nitrogen, water vapor, carbon dioxide, and small concentrations of oxygen and sulfur dioxide.

Particulate carried by the kiln gases originates from the abrasion of the charge as it tumbles through the kiln, the release of particulate due to the gas release associated with calcination, the ash from the fuel if the kiln is coal fired, and the fume resulting from the vaporization and condensation of alkali. The particles resulting from the mechanical abrasion are generally large in comparison with those produced by alkali condensation.

Resistivity of cement-kiln dust is dependent upon moisture content and temperature. In the range of 500°F and above, resistivities are generally in the range suitable for collection by electrostatic precipitators regardless of moisture. However, from 300-400°F, resistivities are extremely dependent on moisture content. In the wet process kiln, moisture is provided by the evaporation of the water from the slurry feed to the kiln. In the dry process, moisture must be provided by evaporation of water in the cooling tower.

Electrostatic precipitators used on cement kilns are of the horizontal-flow, duct-type with insulated steel shells. The application of precipitators to control of cement kiln dust has steadily increased over the past 50 years; the present installed capacity being around 40 million acfm. The trend in the application of precipitators is toward a higher collection efficiency; the average for the past five years being designed for around 99.7%.

Inlet dust loadings for cement kiln precipitators vary from a minimum of around 3 gr/scfd to around 50 gr/scfd. Gas velocities range from around 3-8 ft/sec. Moisture contents range from 4-14% for dry process kilns to 13-40% for wet process kilns. Inlet gas temperatures range from

350-650°F on wet process kilns to 500-700°F on dry process kilns. Precipitation rate parameters for cement kilns range from around 0.25-0.45 ft/sec.

Precipitator applications to cement kilns have been particularly favorable from the standpoint of recovery of cement as well as air pollution control. The effluent dust from the cement kiln has about the same composition as the kiln charge. Consequently, its recovery is of direct economic importance. However, alkali present in the dust cannot be recycled. A fundamental property of the electrostatic precipitator is that larger particles tend to be separated first, and the smaller, alkali-containing fraction of the dust tends to be removed in the last stages. This fractionating effect has been used to separate the dust that is relatively free from alkali and to recycle it through the kiln.

There are several problems associated with the use of electrostatic precipitators in cement kiln applications. (1) In the dry process kilns, the moisture content is low, and consequently resistivity would tend to be high in the temperature range of 300-400°F. There is a trend toward the use of fabric filters on dry process kilns in this country. (2) Upsets in kiln operation can create conditions under which combustible gas can be introduced in the precipitator. Instances of fire in the precipitator have been reported which limit acceptance of precipitators, although this can also be a problem in other types of dust control equipment. (3) Operation at certain temperatures can result in a deposit on the electrodes that impairs precipitator performance.

Gypsum. Gypsum is a hydrated calcium sulfate which, when heated, looses water to form plaster. Removal of the water occurs in a process called cooking. Electrostatic precipitators are used to control emissions from gypsum kettles.

Particulate emissions from the cooking process result from the calcination process and the mechanical agitation of the charge. The size of the dust particles is relatively large compared with processes where the vaporization and condensation of material occur.

The process of cooking increases moisture of the exit gases. The wet gas is mixed with ventilating air and flue gas to give a mixture of gas that ranges from 300-400° F with a moisture content of 30-35%. As a result of the moisture, resistivity of the dust is not usually a problem.

However, at times, water sprays are used in the flue ahead of the precipitator.

The use of electrostatic precipitators in the production of gypsum has steadily increased from around 1935 to the present. The total installed capacity is approximately 1.8 million acfm. The gas flow ranges from a minimum of around 3000 acfm to 14 acfm for kettles to a maximum of around 80,000 acfm for rotary calciners. Efficiencies are generally in the range of 95-99+%. Gas velocities range from 1.5-8 ft/sec. Inlet dust loadings are from around 4-60 gr/scfd. Inlet gas temperatures are around 200-350°F for calciners and 125-250°F for rock dryers. Design precipitation rate parameters for precipitators used in the gypsum industry are around 0.4 - 0.5 ft/sec.

CHEMICAL INDUSTRY

Precipitators are used in the chemical industry in the production of sulfuric and phosphoric acids.

Sulfuric acid. Sulfuric acid is made by the oxidation of sulfur dioxide to sulfur trioxide and subsequent absorption in a recirculating sulfuric acid solution to make acid of the desired concentration. About 95% of the sulfuric acid produced in this country is by the contact process in which SO_2 is catalytically oxidized by atmospheric oxygen in the presence of a vanadium pentoxide catalyst. Sources of SO_2 for the process are burning of elemental sulfur, roaster gases from metallurgical operations, and burning of hydrogen sulfide and spent acid from petroleum refineries.

Electrostatic precipitators can be used in the manufacture of sulfuric acid in two ways. If the source of SO_2 is smelter gas from nonferrous metallurgical operations, the gases contain approximately 3-10% SO_2 and are contaminated with dust, which must be removed prior to being introduced to the converter to prevent fouling of the catalyst. Cleaning can be accomplished by electrostatic precipitators or wet scrubbers.

In the converter, SO_2 present in the gases is converted to SO_3. Gases from the converter exit at around 450°F and pass to the absorber where the SO_3 combines with water to produce 98-99% sulfuric acid.

Gases leaving the absorber contain unabsorbed SO_3 and H_2SO_4 mist. The mist is of very small size and will pass through the absorber without being collected and would be emitted from the process unless suitable collection equipment is provided.

The type of collection equipment needed to remove the acid mist depends upon the size of the particles. Wire mesh eliminators have collection efficiencies of over 90% when most of the particles are greater than 3 microns, which is the case when 98% acid is being produced. When oleum is also produced, 85-95% of the particles leaving the oleum tower is less than 2 microns and wire mesh pads may not effectively remove these fine particles. Electrostatic precipitators have been extensively used for acid mist removal and are effective for the smaller sizes.

Precipitators for acid mist collection have historically been of the vertical up-flow type with cylindrical shells constructed of sheet lead supported by steel banding. Discharge electrodes are of lead with a star cross section. Acid mist precipitators with all-steel construction have been installed in recent years without apparent difficulty.

Between 1945-1969, about 120 sulfuric acid mist precipitators were installed with a total capacity of 2, 230, 000 cfm. The average precipitator size has increased from 10, 000 acfm in 1945 to around 30, 000 acfm in the 1964-1968 period. Design efficiencies are in the range of 97. 5 - 98. 5%.

Gas velocities for acid mist precipitators range from 2-8 ft/sec, with the majority falling in the 3-5 ft/sec range. Inlet loadings range from about 0. 2 - 2. 5 gr/scf. Operating temperatures range from 80-180° F. Input power varies from around 150-700 watts/1000 cfm for the majority of installations. Design field strengths are from 8. 5 - 13 kV/in. Design precipitation rate parameters are around 0. 2 - 0. 3 ft/sec.

Phosphorus. Precipitators are used in the production of phosphorus and phosphoric acid. In the production of phosphorus, phosphate rock, silica and coke are charged in an electric furnace and heated to around 2300-2700° F to liberate P_2O_5, which is reduced to elemental phosphorus by the carbon. The effluent from the furnace is cleaned by an electrostatic precipitator operating above the condensation temperature of the phosphorus (525-620°F) to prevent dust contamination of the phosphorus as it is condensed. Precipitators typically remove 90-99% of the dust.

Precipitators for use in cleaning phosphorus furnace gas are typically vertical gas flow, single-stage types with cylindrical collection electrodes. The collection electrodes are heated to avoid phosphorus condensation and special rappers are often used to avoid damage during rapping.

Special types of hoppers are used for dust removal since the collected dust contains some absorbed elemental phosphorus which ignites on exposure to air.

During 1938-1969, about 20 precipitators have been installed for hot phosphorus applications. The total gas volume handled is around 165, 000 acfm. Gas velocities for phosphorus precipitators range from 1-6 ft/sec, with the average around 2-3 ft/sec. Average gas temperatures range from 500-600°F, with some as high as 800°F. Dust loadings range from 4-15 gr/scf, with the majority in the range of 12-14 gr/scf.

Predictions for phosphorus furnace applications are for continued growth, although the total volume is small in comparison with other areas.

Precipitators for use in the production of phosphoric acid are for mist elimination. During 1930-1960, the cumulative installed capacity was around 150, 000 acfm.

The corrosive nature of the gas creates some problems in the choice of materials. Stainless steel pipes of 5-15 inch diameter are used as collection electrodes. The collected mist drains from the plates so that no rapping is required.

Inlet gas velocities for phosphoric acid mist precipitators range from 2-8 ft/sec. Inlet gas temperatures vary from 150-300° F, and inlet concentrations range from around 5-35 gr/scfd.

Carbon black. Precipitators used in the collection of carbon black are for the purpose of agglomeration of the particles so they can be collected in a mechanical collector. About 18% of the carbon black is collected in the precipitator and 72% in the mechanical collector. Bag filters are often used following the mechanical collectors to reduce the emissions to the atmosphere.

Carbon black particles are extremely fine, ranging from 0.02 - 0.4 micron. The conductivity is also very high, so that the particles would tend to be discharged upon contact with the collection electrode. However, the carbon black particles cling tenaciously to both the discharge and collection electrodes and cleaning is a problem

It is estimated that about one hundred sets of electrostatic precipitator - mechanical collector units have been built since 1926. No information was found on installations made since 1958.

MUNICIPAL INCINERATORS

Use of electrostatic precipitators for control of municipal incinerator emissions is a relatively new application in this country, although the practice is rather widespread in Europe.

There are two principal types of furnaces in general use for incineration of municipal wastes — these are water-cooled and refractory lined furnaces. The type of refuse handled is highly variable between countries and between different sections of the same country. This leads to large variations in the properties and composition of the particulate.

Particle size of the fly ash from municipal incinerators ranges from a mass median diameter of 15 to 30μ. Resistivity of the ash varies with temperature and moisture content and with particle size.

Gases from municipal incinerators are at temperatures in the neighborhood of 1200-1800°F and must be cooled before entering the precipitator. Cooling can be provided by evaporation of water in a water cooling tower or by heat exchangers in a system utilizing heat recovery.

Design precipitation rate parameters for a group of precipitators installed on European incinerators range from around 0.2 - 0.4 ft/sec. Gas velocities range from 2-4 ft/sec. Power densities range from 50-200 watts/1000 cfm.

Precipitation rate parameters for European precipitators vary with precipitator inlet gas temperature. Data from the U.S. installation, for which information was available, agree well with those data from the European installations.

No test data are available on the municipal incinerators in this country since they have been installed so recently.

PETROLEUM INDUSTRY

The principal uses of electrostatic precipitators in the petroleum industry are in the collection of particulate emissions from fluidized bed catalytic cracking units and the removal of tar from various gas streams, such as fuel gases, acetylene, and shale oil distillation gases. The first of these areas, recovery of catalyst dust, originated with the production of high octane gasoline during World War II. Electrostatic precipitators were used to recover catalyst from the discharge stream of the catalyst regenerators as a part of the process. Improvements in mechanical collectors inside the regenerators have eliminated the process requirements. However, precipitators are presently used for control of dust emissions from the process.

Gases from the catalyst regenerator are exhausted from the top through a series of mechanical collectors which return all but the fine particles to the process. The regenerator may be followed by waste heat boilers that recover some of the energy and reduce the temperature of the gases to the precipitator to around 600-700°F.

The inlet dust concentration varies with the type of mechanical collector. A typical range is 0.2 - 1.0 gr/scfd. Mass median particle size is around 10-12 microns. Resistivity varies with temperature and moisture content. Conditioning of the catalyst dust by additions of ammonia to the effluent has been reported to reduce resistivity and improve performance in instances where high resistivity dust has been a problem.

Detarring. Precipitators used for detarring of gases are commonly of the single-stage, vertical wire and pipe type, with collection electrodes suspended from a top heater. The collected oils and tars are usually free flowing and no rapping is required for removal.

During the period 1940-1963, approximately 55 precipitators were installed for detarring of carburetted water gas, 3 for detarring oil gas, 3 for reformed gas, 2 for shale oil, and 1 for acetylene. The total gas volume for all of these applications is around 400,000 acfm. Gas temperatures for detarring precipitators range from around 70-120°F and collection efficiencies are around 95%.

The total market for detarring precipitators is small since very little manufactured gas is sold in this country and the requirements for shale oil processing and acetylene manufacturing are relatively small.

NONFERROUS METALS

The commercial use of electrostatic precipitators has been standard practice by copper, lead, and zinc smelters in cleaning the off-gases from the extraction process. Precipitators are also used in cleaning gases from electrolytic cells in the reduction of bauxite to produce aluminum.

Extraction of nonferrous metals from their ores is carried out in a number of types of processing equipment, many of which are common to all nonferrous operations.

Sinter machines are used to convert metallic ores, fines, and plant process dust into larger material that can be handled in the reduction process. Pellets to be sintered are spread on grates that move the material through the sinter machine. Gas-fired burners ignite the material and air is supplied by fans to maintain combustion. Exit gases are collected in windboxes and passed to the electrostatic precipitator for cleaning.

Ore roasting can be accomplished in a variety of types of furnaces including multiple hearth, flash roasters, and fluid bed roasters. Smelting and refining operations are carried out in reverberatory furnaces and blast furnaces. Converters are used to convert matte to metallic copper. Cupolas are used in the nonferrous metals industry to melt and reduce copper brasses, bronzes, and lead.

Aluminum is produced by the electrolytic reduction of alumina (Al_2O_3) dissolved in a molten cryolite bath. During reduction, electrolytic, thermal, and chemical action in the cell results in the evolution of carbon and alumina dust, other particulates, and gaseous fluorides. Hoods over and around the cells collect the effluent which is sent to mechanical collectors followed by an electrostatic precipitator and then to a scrubber for removal of the remaining gaseous fluorides.

In copper production, the copper-bearing ores are roasted to eliminate some of the sulfur from the concentrate and to volatilize zinc, arsenic, and antimony present in the ores. The roasted ores are then smelted to produce a molten sulfide of iron and copper. The copper matte from the smelter is then converted to copper by blowing it with air in a converter. The blister copper from the converter is then refined by either fire or electrolytic methods.

Lead production begins with either oxide or sulfide ores. Oxide ores can be directly reduced in a blast furnace, whereas sulfide ores must be first converted to oxides. This is accomplished by roasting or sintering in an oxidizing atmosphere. After converting to the oxide, the lead ore is reduced in a blast furnace by reaction with carbon supplied by coke charged in the furnace. Refining of the lead is accomplished by electrolytic refining or by a kettle or reverberatory furnace.

Zinc. Zinc is generally extracted from ores containing both zinc and lead sulfides, although some zinc and copper-zinc ores are processed. The zinc ores are concentrated by flotation and processed in a roaster to convert the zinc sulfides to zinc oxide. Metallic zinc is produced from the roasted ore by retorting, electrolysis or fractional distillation.

Most of the electrostatic precipitators used in the nonferrous metals production have been designed by the large western smelters, and hence information on the applications in the nonferrous metals industry is limited. Information that is available indicates that design precipitation rate parameters vary from 0.05 - 0.07 ft/sec for precipitators operating on converter gas, 0.12 - 0.14 ft/sec for precipitators used on copper roasters and reverberatory furnaces, and around 0.25 ft/sec for sinter machine precipitators.

HIGH TEMPERATURE, HIGH PRESSURE GAS CLEANING

A unique application of precipitators would be in the cleaning of high temperature, high pressure gases. One of the early applications of precipitators was in the removal of ash from the products of high temperature gasification of coal. This work was a pilot-scale research operation conducted in conjunction with the development of the coal-fired, gas-turbine locomotive. Temperatures of 1500°F and pressures up to 600 lb/in.^2 were used in this study.

Current interest in high temperature, high pressure gas cleaning is for the use of the gases produced from waste incineration to produce electric power from gas-turbine-driven generators.

Research has been conducted on corona generation at temperatures up to 1700°F and pressures of 100 psig. Precipitation rate parameters of 0.23 - 0.26 ft/sec were obtained on pilot-scale units and extrapolations made to full-scale precipitators.

NEW APPLICATION AREAS

New applications for electrostatic precipitators can take the form of replacement of other types of dust control systems, or the control of particulate emissions from sources where no particulate control devices are now used.

The major advantages of electrostatic precipitators are that high collection efficiencies can be achieved even with small particles and the pressure drop across the precipitator is low. The latter characteristic makes precipitators especially attractive when large gas volumes are to be handled.

The high initial cost of electrostatic precipitators is a disadvantage. However, when determining air pollution control costs, total costs over a period of years should be determined.

Two factors that limit application of electrostatic precipitators are: (1) high resistivity dust results in limitation of the operating voltage and current so that the resulting precipitation rate parameter is low. This necessitates the use of an excessively large precipitator, alterations in inlet gas temperature, or additions of chemical conditioning agents to alter dust resistivity. These constitute additional costs and can make other dust control methods more attractive. (2) Very fine particles do not acquire a charge sufficient for good precipitation. This again results in low precipitation rate parameters requiring larger precipitators with higher costs.

Two primary areas for increased precipitator applications are in the control of emissions from municipal incinerators and in control of foundry cupola emissions. The use of precipitators on municipal incinerators is relatively common practice in Europe. Within the past year, several installations have been made in this country.

The control of cupola emissions by electrostatic precipitators is potentially a promising area. However, details of accommodating the highly variable emission rates and character of the emissions must be resolved.

A summary of the use of the various types of dust control equipment in each of the areas identified by SIC classification is given for the 1966-1967 period. These represent potential use areas for electrostatic precipitators.

CHAPTER 15
ELECTROSTATIC PRECIPITATORS IN THE ELECTRIC UTILITY INDUSTRY

Electrostatic precipitators are used in the electric utility industry for the collection of fly ash from fossil fuel fired boilers. This application constitutes the largest single use of electrostatic precipitators in this country.

Sources of energy for electric power generation are coal, oil, natural gas, hydroelectric, and nuclear. In 1937, approximately 55% of the energy for electric power generation was supplied by combustion of coal.[1] Over the past 30 years, the percentage of the total energy supplied by coal combustion has steadily declined. Estimates of the percentage of energy sources for power generation given in Table 15.1 show that coal will supply approximately 18-22% of the total energy by 1980. However, during the period from 1937 to 1980, the total energy used for electric power generation will have increased by a factor of approximately 3.9. Thus, the total energy supplied by coal is expected to increase.

The decrease in the percentage of energy supplied by coal has been largely due to increases in the use of oil and natural gas. Nuclear power is estimated to supply about 5-9% of the energy by 1980. However, delays have been experienced in nuclear plant construction due to site location controversy, delays in development of breeder reactors, and other factors, so that there is some uncertainty in the schedule for nuclear plants.

Of the known reserves of recoverable energy in this country, about 83% is estimated to be in the form of coal. Much of this coal reserve is in the large Western coal regions and contains less than 1% sulfur. Because of these large reserves, the use of coal as an energy source for electric power generation is expected to continue for the foreseeable future with some new coal burning plants being added to supply the increased energy requirements.

The total quantity of fly ash emitted from heat and electric power generation before collection is given in Table 15.2.[2]

The modern coal fired electric generating plant is comprised of a boiler, generator, condenser, coal handling equipment, dust collection and

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[1] Refer to the bibliography for this chapter.

Table 15.1

Energy Sources for Electric Power Generating Plants[1]

	1937	1947	1957	1965	1980
Total Energy 1937 = 100%	100	144	184	237	389
Coal	55	48	27	23	18-22
Oil	30	34	44	43	41
Natural Gas	11	14	25	30	28
Hydroelectric	4	4	4	4	4
Nuclear	-	-	-	0.1	5-9

Table 15.2

Particulate Emission from Heat and Power Generation before Collection (Millions of Tons per Year)

			Particulate Emission		
Source	Type Firing	1940	1960	1980	2000
Utilities	Pulverized coal	4.9	12.34	24.20	27.1
	Stoker coal	-	-	-	-
	Cyclone coal	-	0.25	0.49	0.55
	Oil	0.005	0.018	0.022	0.024
Industrial	Pulverized coal	1.43	1.87	2.12	2.61
	Stoker coal	3.80	3.60	4.09	5.05
	Cyclone coal	-	0.05	0.06	0.07
	Oil	0.038	0.054	0.083	0.157
Residential and Commercial	Pulverized coal	-	-	-	-
	Stoker coal	3.85	1.98	0.91	0.35
	Cyclone coal	-	-	-	-
	Oil	0.150	0.159	0.169	0.197
Total		14.173	20.331	32.144	36.108

disposal equipment, water handling and treatment facilities, and heat recovery systems such as economizers and air heaters. Boilers in present day usage are of the cyclone and pulverized fuel types, with pulverized fuel boilers comprising approximately 90% of the total. Stoker fired boilers were used in earlier plants, but have been largely outmoded.

Pulverized fuel boilers utilize coal ground to a size such that about 70% passes a 200-mesh screen. The coal and preheated combustion air are direct-fired into the boiler as determined by the steam requirements. The coal-air ratio is automatically regulated to give optimum combustion for all load conditions. Boilers are classified as horizontal, vertical, or tangential, depending upon the firing position of the burners. Primary, secondary, and in some cases tertiary air is introduced in a variety of ways, depending upon the design of the particular boiler manufacturer.

Pulverized coal fired boilers are also classified as either wet bottom or dry bottom, depending on the operating temperature and ash fusion temperature. In wet bottom boilers, the temperature is maintained above the ash fusion temperature so that the slag is molten and can be removed from the bottom as a liquid. Dry bottom boilers operate at temperatures below the ash fusion point, and the ash is removed in a solid state.

Cyclone boilers operate with much coarser fuel, typically 95% minus 4 mesh. The heater is a water cooled cylinder with combustion air introduced tangentially. Combustion occurs at high heat release rates (500,000 to 900,000 Btu/cu ft-hr) at temperatures sufficient to melt a high percentage (85 to 95%)[3] of the ash which is discharged through slag tap openings.

15.1 FLY ASH CHARACTERISTICS

Concentration. Both pulverized coal and cyclone boilers generate considerable quantities of ash, part of which is carried out with the combustion gases. The amount of fly ash depends on several factors, the more significant ones being composition of the coal, boiler design, and boiler steam rate. Other conditions remaining fixed, fly ash emission will be approximately proportional to the ash content of the coal. Boiler design and operation determine the percentages of ash retained in the furnace and emitted in the flue gas. Boilers for high-fusion ash are usually of the dry bottom types which retain approximately 20% of the ash, while wet bottom boilers used for low-fusion ash may retain as much as 40% of the ash. As previously mentioned, cyclone boilers may retain 85 to 95% of the ash initially contained in the coal. Coals encountered in practice have ash contents ranging between 5 and 25%, with an average value of around 8 to 15%. The range of ash content for coals burned in a group of 48 power plant boilers is shown in Figure 15.1. The fly ash concentration in the flue gas ranges from about 1 to 7 gr/cu ft, with an average value of about 3 gr/cu ft.

Particle size. Due primarily to the direct relationship between particle size and particle charge, size distribution is an important factor in electrostatic precipitation. In addition to its effect on particle charge, particle size may be a significant factor in reentrainment and in the electrical properties of the collected dust layer.

Reported particle size distribution measurements from several installations indicate that boiler furnace design and operating conditioning are the most important factors in determining fly ash particle size.[4] Fly ash particles from pulverized coal furnaces are typically larger than those from cyclone type furnaces. In cyclone boilers, it is common practice to recycle the fly ash to the boiler, which results in a greater percentage of fines due to attrition of the fly ash particles. There is also a rather large variation in particle size for each type of furnace, which may be partially attributed to variations in coal grinding and boiler operation conditions, but is probably also influenced by coal composition. While no quantitative relationships have been established, it has been empirically observed that the coarser ash tends to be associated with coal having a high combustible content, and the finer ash with low combustible content coal.

In some power plant fly ash collection systems, mechanical collectors precede the electrostatic precipitators and serve to reduce the particulate load to the precipitator. In addition to reducing the fly ash concentration, these mechanical collectors (generally of the multiple cyclone type) remove a significant portion of the size fraction larger than 10 to 20μ diameter. Typical collection efficiencies for mechanical collectors are in the range of 65-75%.

As seen in Figure 15.2, the size distribution for pulverized fuel furnaces is typically shifted toward a predominantly smaller size when the electrostatic precipitator is preceded by mechanical collectors. The

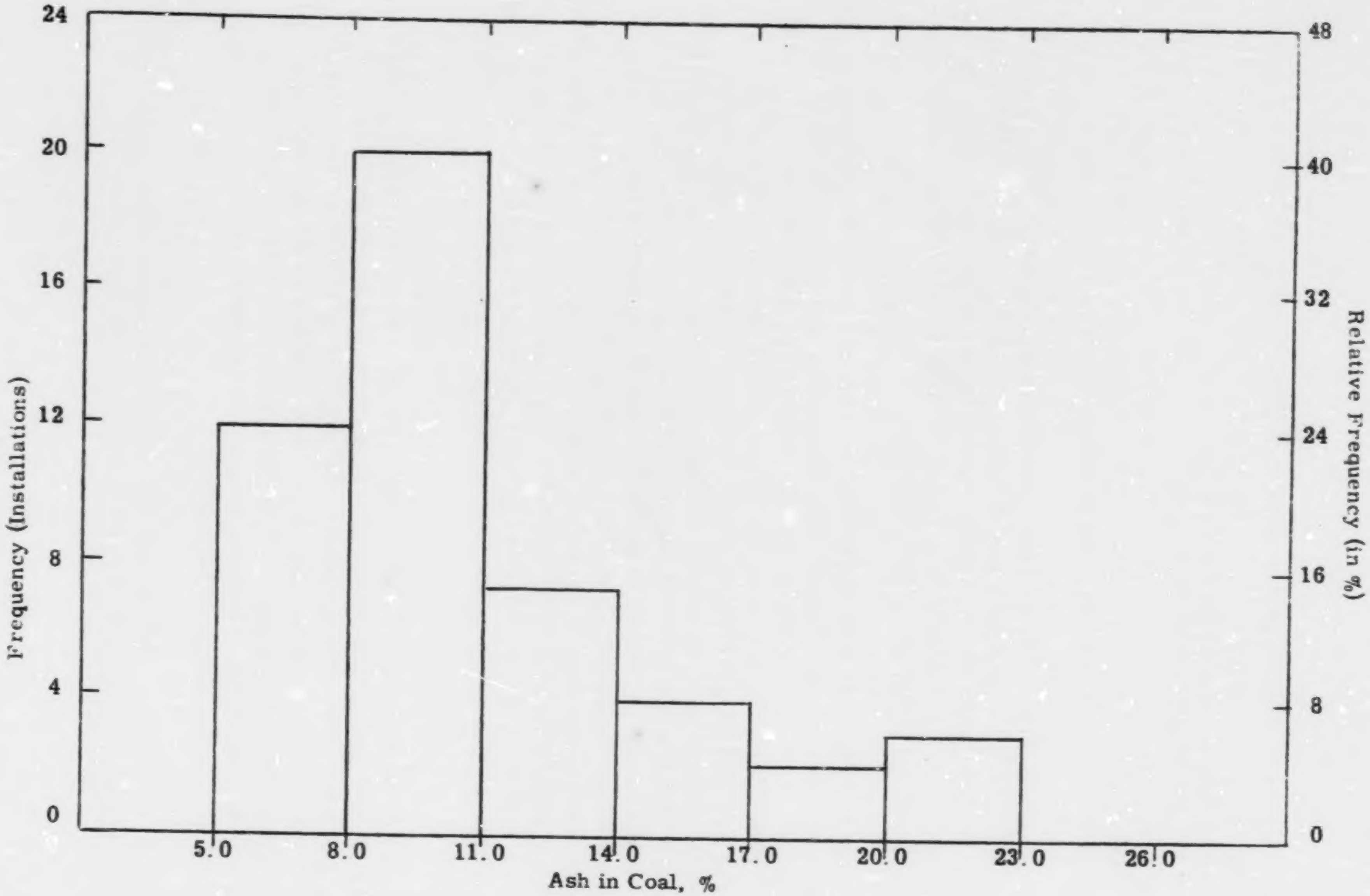


Figure 15.1. Frequency Distribution of Percent Ash in Coal Burned in 48 Different Power Plant Boilers on Which Precipitators are Installed.

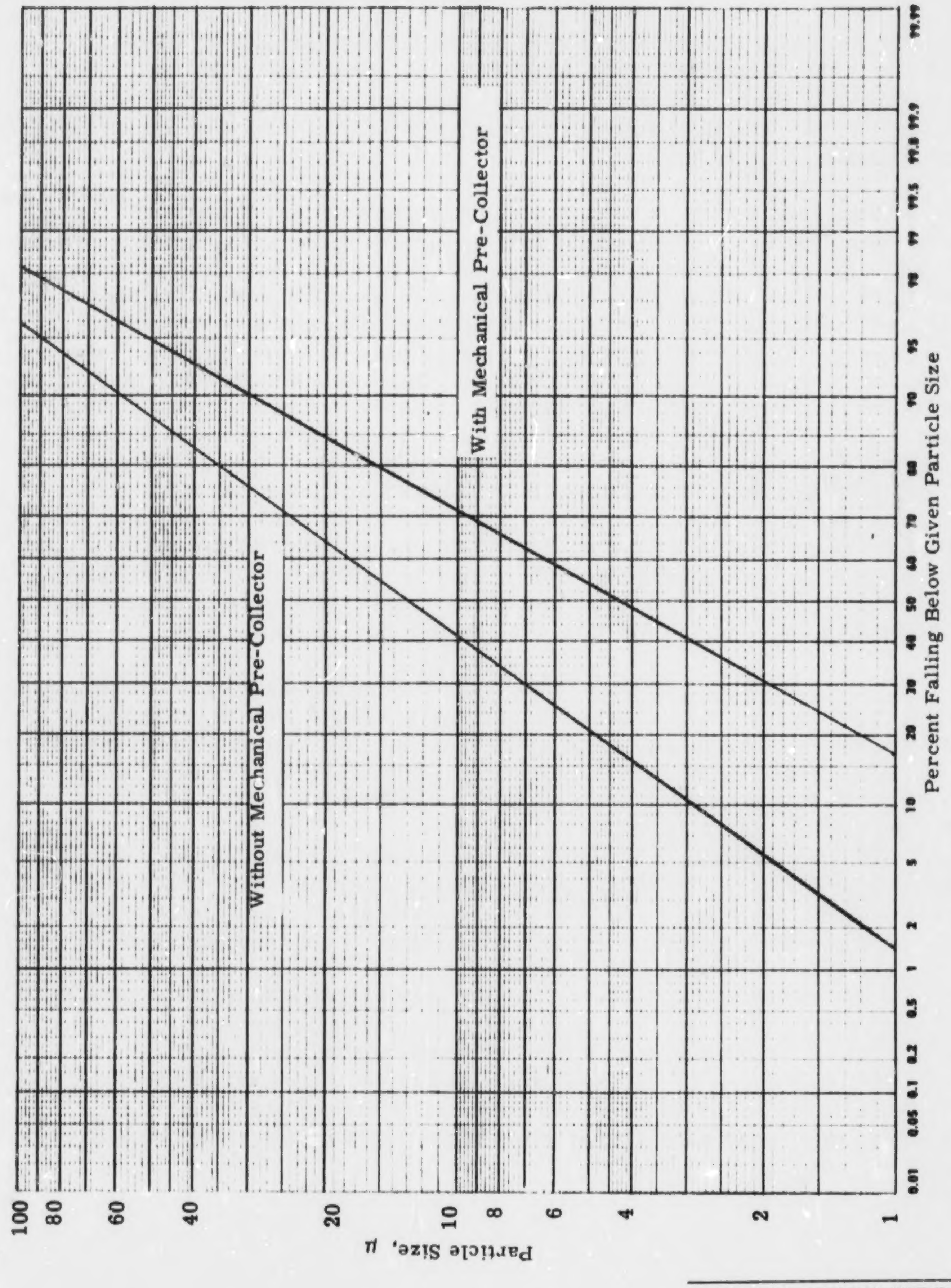


Figure 15.2. Typical Fly Ash Particle Size Distribution for Pulverized Fuel Furnaces.[4]

fly ash particles are approximately log-normally distributed, with a geometric mean of approximately 15 μ without mechanical collectors, and a geometric mean of approximately 4 μ diameter with mechanical collectors installed. Figure 15.3 shows a typical particle size distribution range for fly ash from cyclone furnaces.

Some consideration should be given to the method of particle size determination in evaluating size distribution of fly ash. One method commonly employed[5] consists of wet screening the $\geq$ 325 mesh fraction through a series of standard screens, and separating the $\leq$ 325 mesh in a centrifugal particle classifier. Another method consists of suspending the ash in an electrically conductive liquid and determining the size-number distribution electronically. In a comparison of these two methods, a significant difference in size distribution has been noted for the same fly ash. The wet screen-classifier method showed fewer large particles than the other method.[5] The centrifugal classifier method is specified by ASME Power Test Code No. 28 for the evaluation of dust separating apparatus, and is widely used in analysis of size distributions. It should be noted, however, that the method is dependent upon particle shape and density factors, and hence is a secondary standard which must be calibrated. Other methods utilized include the use of light and electron microscopy, sedimentation, and elutriation methods for size-distribution determinations.

Physical appearance. Individual fly ash particles range in size from very fine to very coarse, depending upon the sample source. Particle colors range from a light tan or gray to black. The tan shades are usually due to the presence of iron oxide; whereas, the dark shades are usually indicative of unburned carbon or magnetic iron oxide.

Microscopic examination of fly ash shows it to be composed of a variety of shapes. There is a predominance of hollow spherical translucent particles, often referred to as cenospheres.[6] They cover a wide range of sizes from less than 1μ to over 300μ. Unburned carbon particles are found in a variety of shapes, with some having a characteristic coke-like microstructure which is commonly referred to as grit. Some fly ashes also contain relatively large white particles that may originate from mineral matter in the coal or, in some cases, from agglomeration of finer ash particles while still in a fused state. The percentages of these constituents present in fly ash can vary widely from sample to sample, depending upon such factors as the composition of the coal

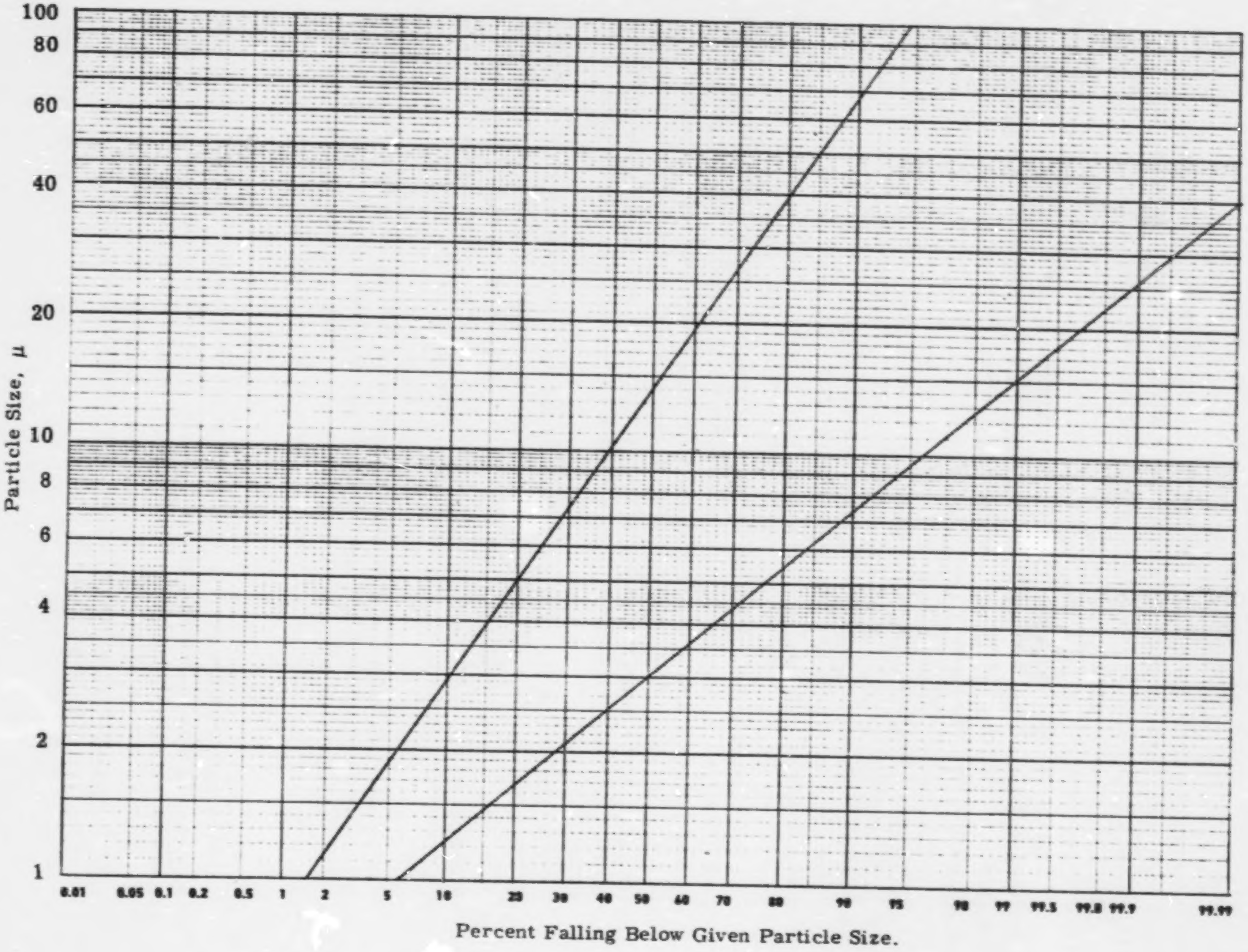


Figure 15.3. Typical Range of Particle Size Distributions for Fly Ash from Cyclone Furnaces.

burned and furnace operation. Because fly ash is such a heterogeneous and highly variable material, it may have widely different physical and chemical properties from plant to plant, and sometimes even from day to day in the same plant.

The presence of carbon in fly ash indicates incomplete combustion in the furnace and a loss of fuel. It is therefore desirable to reduce the carbon lost with the ash to the lowest practicable amount consistent with a favorable air-fuel ratio.

Density. The bulk density of a typical fly ash ranges from 30 to 50 lb/cu ft (0.5 to 0.8 g/cm^3), although much higher densities can be found. The density of each size fraction will vary considerably, with the relatively large coarse particles having the lower density, typically of the order of 0.6 to 1.0 g/cm^3.

The finer particles tend to be lower in carbon content and have a much higher density, usually in the range of 1.5 to 3.0 g/cm^3. The large difference in density and structure of the fine and gritty particles frequently tends to produce a stratification or separation of the particles in the gas. The large, gritty particles have very low adhesive and cohesive characteristics, and are, therefore, easily redispersed into the gas stream.

The physical state of the fly ash is variable, depending on its relative age and temperature. Freshly precipitated hot, dry ash flows somewhat as a liquid. Cold, damp ash, on the other hand, has a tendency to cling tenaciously to electrode and hopper surfaces, and can form a mass in the hopper which is difficult to remove (concreting).

Chemical composition. Fly ash is composed chiefly of silicates, oxides, sulfates, and carbon. In addition to these main constituents, a large number of elements may be present in minute quantities. The composition, or relative percentages, of these constituents covers a wide range for different ashes.

A comprehensive source of information on the composition of fly ash is a U. S. Bureau of Mines summary of the chemical analyses of over 300 samples of fly ash. Included in the report is information on the coal from which the ash was produced.[7] Information on the fundamental characteristics of fly ash from combustion of pulverized coal was also determined by the ASTM.[8] This report, which includes data on 39 fly ash samples, indicates large variances in the chemical and physical properties of the fly ash samples. X-ray diffraction tests show the presence of some crystalline material, magnetite, and possibly mullite. The mean ranges of the major constituents of the low-and medium-carbon fly ashes are as follows:

Constituent	Weight, %
SiO_2	42-50
Al_2O_3	20-30
Fe_2O_3	12-23
CaO	2-7
MgO	0.5-1.5
SO_3	0.1-2
Available alkali	0.4-1.5
Carbon	2-20

Ash-softening temperature. The ash-softening or fusion temperature is important to the character of the fly ash. Fly ash having a low fusion temperature will tend to form glassy spherical particles that are different from other fly ash particles in density and surface properties. These properties can materially affect the resistivity of the fly ash. A coal high in pyrite sulfur (Fe_2S) gives fly ash that is high in iron content, which tends to lower the fusion temperature of the ash. For example, coals having sulfur contents of 3 to 10% have ash fusion temperatures of about 2000 to 2400° F, but coals having about 1% sulfur have ash fusion temperatures of 2600 to 3000° F. The ratio of acidic components (SiO_2) to basic components (Al_2O_3, MgO, CaO) of the fly ash affects the fusion temperature as well as the resistivity of the fly ash.

Resistivity. The heterogeneous nature of the coal burned in electric power generating plants and the variations in operating conditions result in fly ashes that have electrical resistivities that vary typically from about 10^8 to 10^{12} ohm-cm. Such variations in resistivity result in widely different requirements for electrostatic precipitators to meet given performance demands. This variation in resistivity, more than any other single property, accounts for the widely varying performance of fly ash precipitators.

The values of resistivity reported in the literature are often given without regard to the method of measurement. Laboratory values are generally higher than those measured in the field, often by several orders of magnitude. Since resistivity of fly ash is dependent on the ambient gas composition and temperature, the only meaningful measurements are those taken in-situ under typical operating conditions.

Since fly ash is composed primarily of silicates and metallic oxides, its inherent bulk resistivity will be high in the range of operating temperatures of most fly ash precipitators. In this temperature range, where surface conduction predominates, the moisture content of the flue gas is apparently too low (5 to 15% by volume) to provide adequate surface conduction in the absence of additional conditioning agents.

It is generally accepted that the surface conduction properties of the fly ash are related to the sulfur content of the coal. The sulfur trioxide formed acts as a secondary conditioning agent which reduces the resistivity of the ash to the range where it can be effectively collected by electrostatic precipitators.

The fraction of sulfur which is converted to SO_3 (1 to 3% of the SO_2 formed) is apparently the most important single factor in determining fly ash resistivity. It is theorized that either the SO_3 combines with moisture in the flue gas and condenses as an aqueous phase with a resultant increase in surface conductivity, or that SO_3 is adsorbed on the surface of the fly ash and tends to promote the adsorption of water and thereby increase surface conductivity. There is evidence to support each of the theories, and further evidence that in some cases, the formation of sulfate salts on the surface may result in increasing resistivity values with time as the reaction progresses. Consequently, one would expect that regardless of the mechanism involved, the acid-base character of the ash is also quite influential in the conditioning phenomenon. Although there is good correlation on a statistical basis between sulfur content of the coal and resistivity of the fly ash, the variations are so large that the resistivity of a given fly ash cannot be predicted with any degree of confidence on the basis of the composition of the coal. The most reliable method of determining resistivity is to measure it in-situ in the specific plant or in plants using similar boilers burning the coal in question.

One problem in the design of precipitators for electric power generating plants is that the source of coal is often not available at the time the specifications are prepared. In many instances, core samples are the only quantities of the fuel on hand, so that field tests are not practical. In these cases, one has to rely on the coal analysis and make an "educated" guess as to the resistivity values. Graphs of the form shown in Figure 15.4 are useful when sufficient data have been collected. However, it should be recognized that these curves represent average resistivity values and that the resistivity can vary over a wide range for a given sulfur content and temperature.

15.2 FLUE GAS CHARACTERISTICS

The chemical composition of flue gas from power plant boilers will vary according to the composition of the coal. The moisture content of the combustion gases will depend somewhat on the moisture content of the coal and air; usually, however, the volatile hydrocarbon content of the coal is of greater significance due to the formation of the combustion products CO_2 and H_2O. Typical moisture contents of flue gas range from 5 to 15%.

The SO_2 content of the flue gas is a function of the sulfur content of the coal. The average sulfur content of coals used for electric power production in the United States is about 2.5%, as shown in Figure 15.5. A typical concentration of SO_2 in flue gas from the combustion of coals with 2.5% sulfur is around 1500 ppm. The amount of SO_2 in the gas, however, may depend upon the form of the sulfur in the coal. Sulfur can appear in combined form as iron pyrite, organic sulfur, or sulfate compounds.

The portion of SO_2 that is converted to SO_3 depends upon many factors including the type of boiler, the amount of excess air used, and the composition of the coal. It is difficult to accurately analyze for SO_3 in a full scale power plant so that information on the SO_3 content is not too well established; however, it is generally accepted that about 1 to 3% of the SO_2 is converted to SO_3. On this basis, the SO_3 content of the flue gas from a power plant burning 2.5% sulfur coal might range from 15 to 45 ppm.

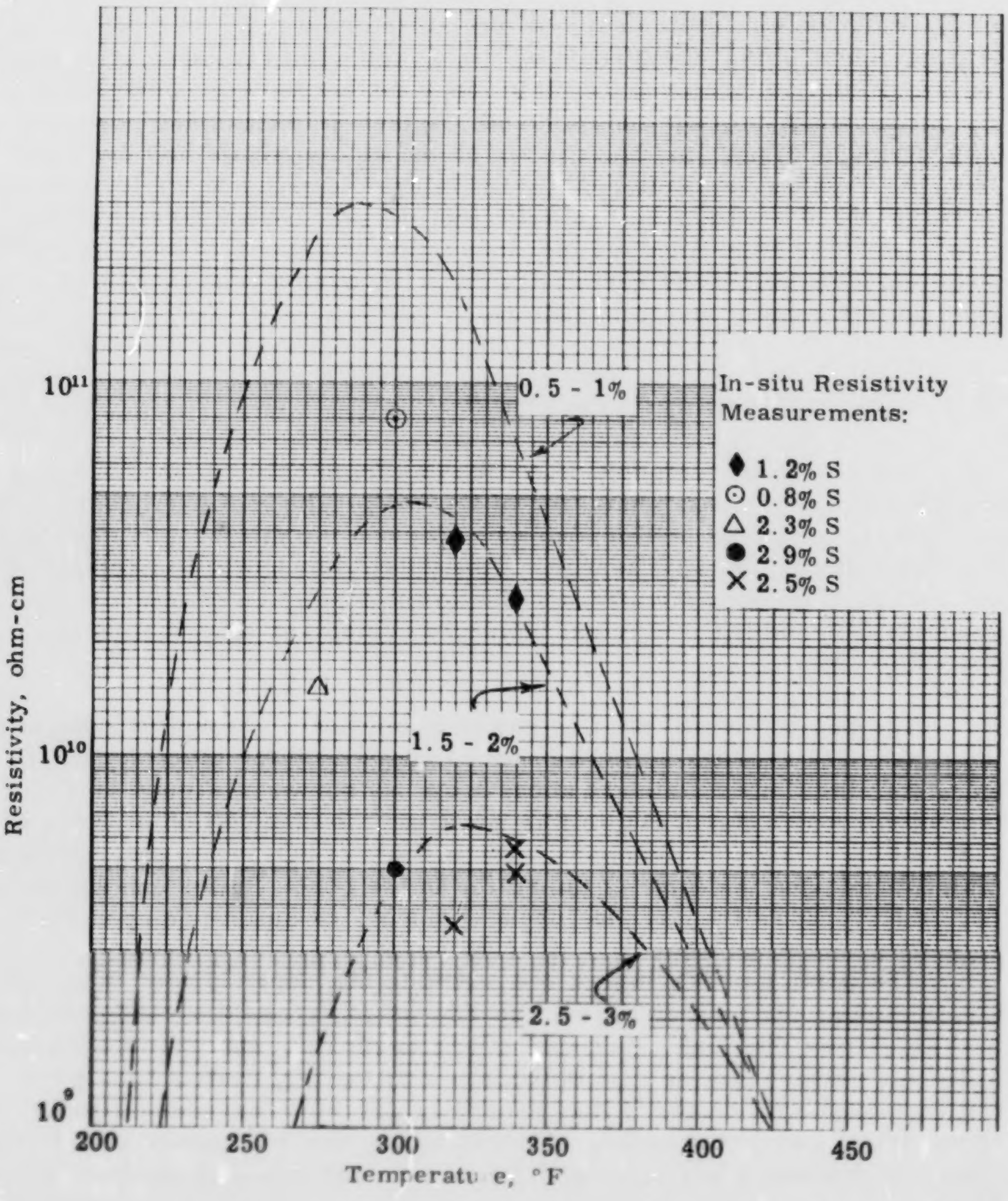


Figure 15.4. Trends in Resistivity of Fly Ash with Variations in Flue Gas Temperature and Coal Sulfur Content.

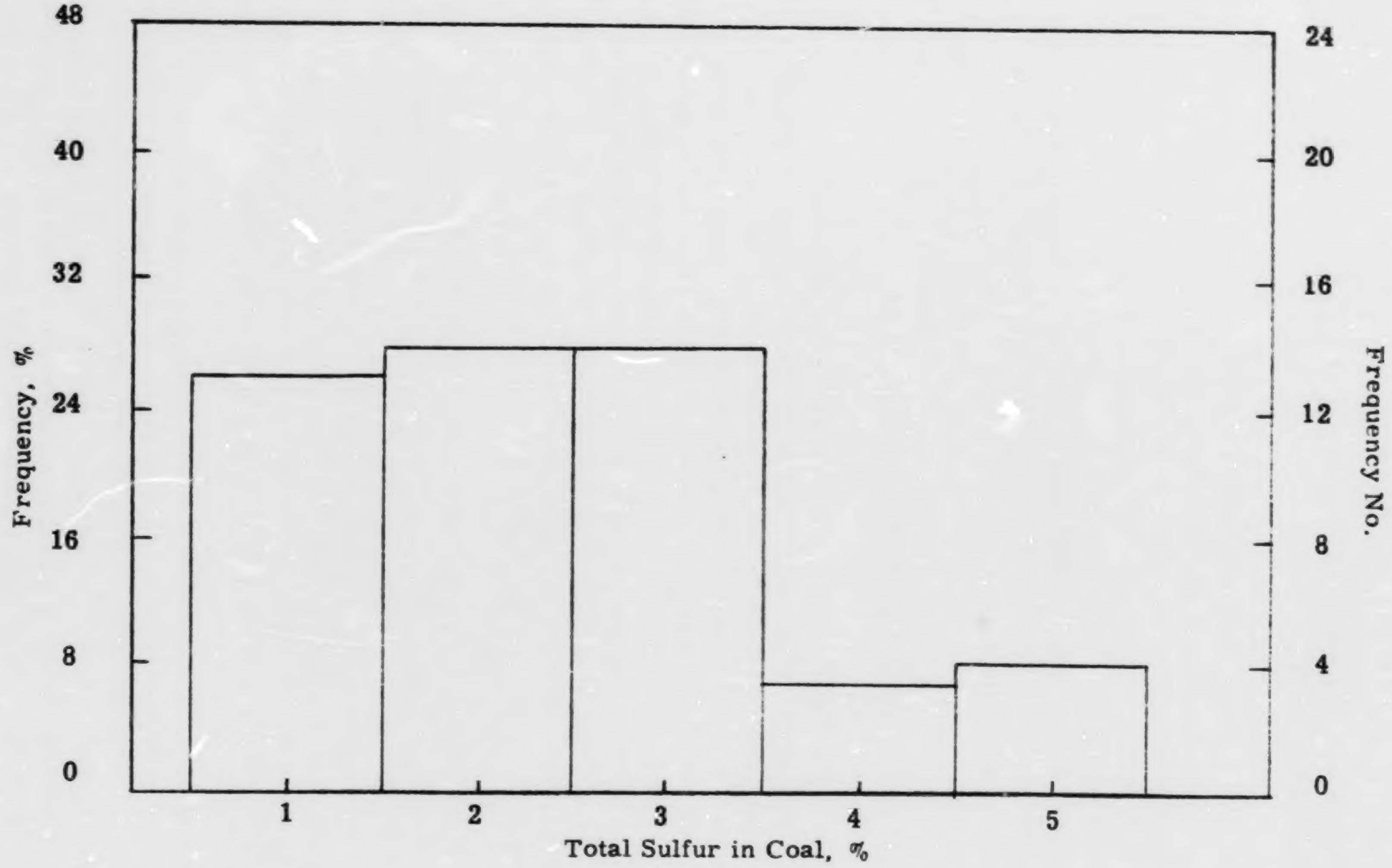


Figure 15.5. Frequency Distribution of Percent Total Sulfur in Coal Burned in 48 Different Power Plant Boilers on Which Precipitators Are Installed.

Typical concentrations of moisture, CO_2, O_2, and excess air used are given in Table 15.3 .

Oxides of nitrogen are formed by combination of nitrogen and oxygen during combustion. The main factors in NO production are: the flame and furnace temperature, the length of time that combustion gases are maintained at flame temperatures, the cooling rate of the gases, and the amount of excess air. The reported concentration of oxides of nitrogen ranges quite widely--100 to 1460 ppm. Since the oxides of nitrogen constitute a potential pollution problem, there is interest in keeping the concentration at a minimum.

15.3 HISTORICAL DEVELOPMENT OF FLY ASH PRECIPITATORS

Historically, the use of electrostatic precipitators for the collection of fly ash from coal fired power plants dates to the early development of modern power generation techniques at the turn of the century. Prior to this time, power plants burned coal in relatively inefficient stoker fired boilers; however, with the development of high capacity, high efficiency pulverized coal fired boilers, fly ash entrainment in the effluent gases became a problem. The increased use of electrostatic precipitators in power plants has paralleled the rapid increase in power production, with the result that over half of the total precipitators installed are for fly ash collection.

The first commercial fly ash precipitators were relatively small units, designed for 90% removal efficiency. These precipitators were of the horizontal flow type, with reinforced concrete collecting electrodes. Dust removal from the collecting electrode was by a pneumatically operated scraping device, which caused considerable reentrainment during the cleaning period. In order to reduce these losses, the cleaning cycle was arranged to coincide with low velocity flow at reduced boiler loads.

The problem of collecting fly ash is becoming increasingly complex due to the increasing boiler sizes and the increasing use of higher ash and low sulfur content coals. In addition, changes in the general practice of base loading newer boiler units, and more stringent air pollution control regulations have resulted in changes in the requirements for collector efficiency. Most precipitators installed today on power plant boilers are designed for efficiencies of $\geq$ 99%.

Table 15.3

Typical Flue Gas Compositions of Selected Components Measured at Moisture Electrostatic Precipitator Inlet[6]

Test No.	Coal Rate (ton/hr)	Flue-Gas Volume scfm	Average Flue-Gas Temp °F	Moisture %	CO_2 %	O_2 %	Excess Air %
1	64.5	394,600	290	6.8	12.6	6.1	39.9
2	65.2	395,800	280	6.1	12.5	6.35	42.2
3	67.0	401,900	365	6.2	12.8	6.2	40.9
4	48.0	287,800	230	5.6	12.9	6.2	41.1
5	46.2	306,700	260	7.2	12.1	7.0	48.9

15.4 DESIGN OF FLY ASH PRECIPITATORS

Two approaches to the specification of fly ash precipitators for a given application are followed. One approach is to specify the type fuel burned, the gas volume to be handled, and the other factors such as are given on the IGCI bid specification form, and to specify a performance either in terms of the maximum outlet dust loading or precipitator efficiency. The manufacturers then determine the area of collecting surface and all precipitator parameters required to meet the specifications. In some instances, however, the user will specify key design parameters based on his own experience. Typically, the specific surface area (ratio of area of collecting plate to gas volume), current per length of discharge electrode, total number of transformer rectifier sets, and the number of rappers per sq ft of collecting surface and per ft of discharge electrode will be specified in the invitation to bid. These parameters are based on the user's experience with particular types of fuel and operating conditions and design on this basis is practical only in the case of large electric utility systems with considerable experience in precipitator operation. In such instances, very little in the way of size is left as an option of the manufacturer.

In the former case, there is often a wide variation in the size precipitator, degree of electrical sectionalization, size of power supply, and other parameters that are bid for a given installation. Hence the user is faced with a considerable problem in evaluating the bids.

Techniques for arriving at the size and electrical requirements for fly ash precipitators are those outlined in the chapter on design methodology in Part I. However, because of the variations in dust properties, the methods of selecting a precipitation rate parameter are less precise for fly ash precipitators than for most other applications.

Precipitator size. The precipitator size required to achieve a given efficiency is determined by the gas flow and the precipitation rate parameter as given by the Deutsch-Anderson equation $\eta = 1 - \exp(-\frac{A}{V} w)$, following method 1 as given in the chapter on design methodology. Gas volume, efficiency, type of fuel, and gas temperature are specified operating conditions. The collecting surface area, A, depends upon the value of the precipitation rate parameter selected.

Figure 15.6 shows the range of precipitation rate parameters for a group of fly ash precipitators. Both design and test values are shown.

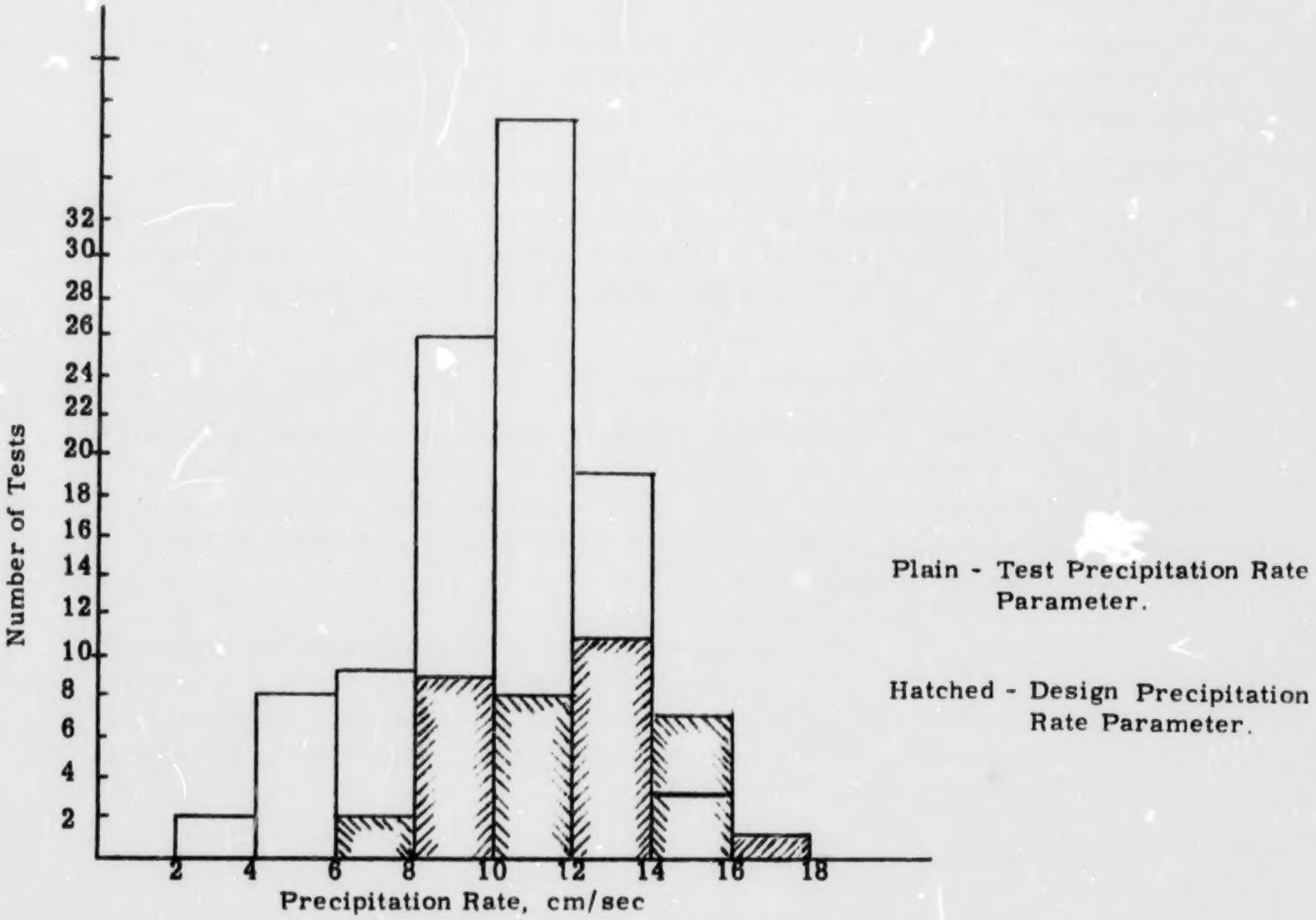


Figure 15.6. Distribution of Precipitation Rates for Fly Ash Precipitators.

As indicated in the histogram, test values of w can vary from around 3 cm/sec (0.1 ft/sec) to around 17 cm/sec (0.56 ft/sec).

Particle resistivity, particle size, gas flow quality, electrical energization, mechanical design, and the state of repair of the precipitator are the principal factors influencing the value of the precipitation rate parameter.

In the design process, there are several methods of arriving at the value of w, and each manufacturer will perhaps use different procedures. In general, the following methods can be used to arrive at design values or to determine the adequacy of a particular design:

Design by analogy

Precipitator manufacturers maintain data on a number of installations and selection of the value of w can be made on the basis of an equivalent operating installation. The accuracy with which w can be selected can be quite good if data are available from a plant burning the same coal in the same type boiler.

The value of w selected on the basis of analogy with an assumed equivalent installation can be adversely influenced by conditions within the precipitator to which the comparison is being made. Alignment of the electrode structure, variations in gas distribution, etc. must be similar for the analogy to be valid.

Design by fuel composition

For new installations, where no equivalent operating installation is available for comparison, selection of the value of w must be made on the basis of available information, which is often limited to an analysis of the composition of the coal.

It has long been recognized that the sulfur content of the fuel is an important parameter influencing the resistivity of fly ash, and sulfur content has been used as a basis for selection of design w. Ramsdell[10] made a study of the performance of a number of precipitators and developed a series of curves relating efficiency and collecting surface area to gas volume ratio with sulfur content as a parameter. Figure 15.7 is a plot of the Ramsdell data.

The data given by Ramsdell can be used to determine the precipitation rate parameter w in the Deutsch-Anderson equation. The values of w are plotted as a function of sulfur content in Figure 15.8. Also shown in the same figure are data taken from installations from other plants to indicate the variations that might be expected.

As indicated previously, sulfur influences w principally by altering the resistivity of the dust. However, factors other than sulfur influence the resistivity. Some coal crushers reject the sulfur appearing as pyrite, and hence the sulfur content of the coal being burned would be different from that entering the plant. Thus the form of the sulfur would result in different resistivities.

The nature of the ash also influences resistivity. As indicated in the chapter on resistivity and conditioning, the rate of adsorption as well as possible reactions with the ash, alter the resistivities that result from burning coal with a given sulfur content. Temperature is also an important parameter in determining resistivity, and the curves shown apply for gas temperatures of 300°F. The surface area of the fly ash is another variable influencing resistivity, and wide variations can result in large changes in resistivity and precipitation rate parameters.

Design by fly ash resistivity

A more fundamental approach to selection of w would be by resistivity of the dust. If resistivity can be measured directly, influences of coal and ash composition, specific surface of the ash, etc., can be eliminated.

Figure 15.9 is a plot of w as a function of resisitivity taken from a number of field measurements on operating precipitators as given by White.[11]

One of the problems associated with design by resistivity is that the dust resistivity value is usually not known on new installations. In some cases, for example, where mechanical collectors are to be replaced by electrostatic precipitators, measurements of the dust resistivity can be made.

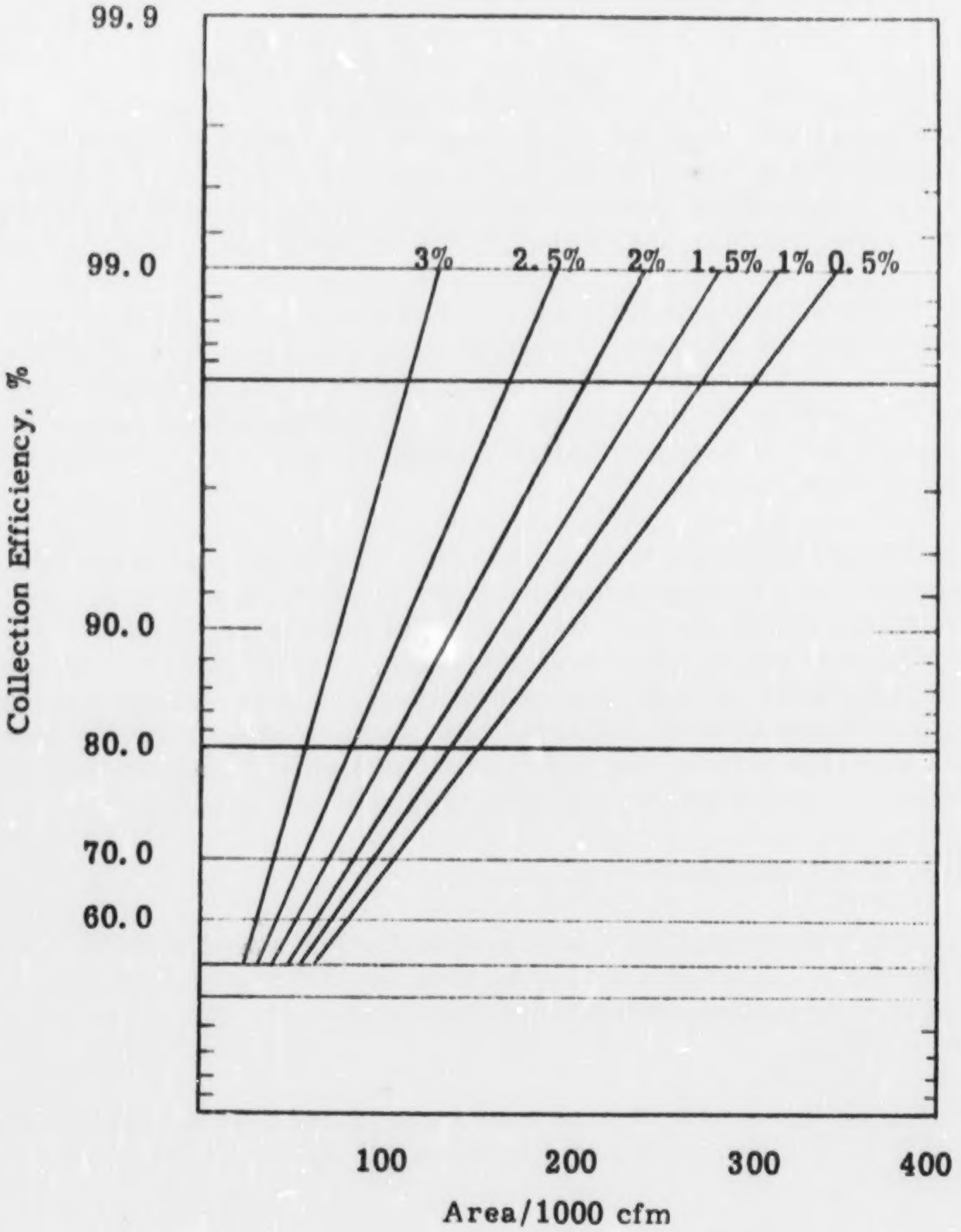


Figure 15.7. Relationship between Collection Efficiency and Collecting Surface Area to Gas Flow Ratio for Various Coal Sulfur Contents.

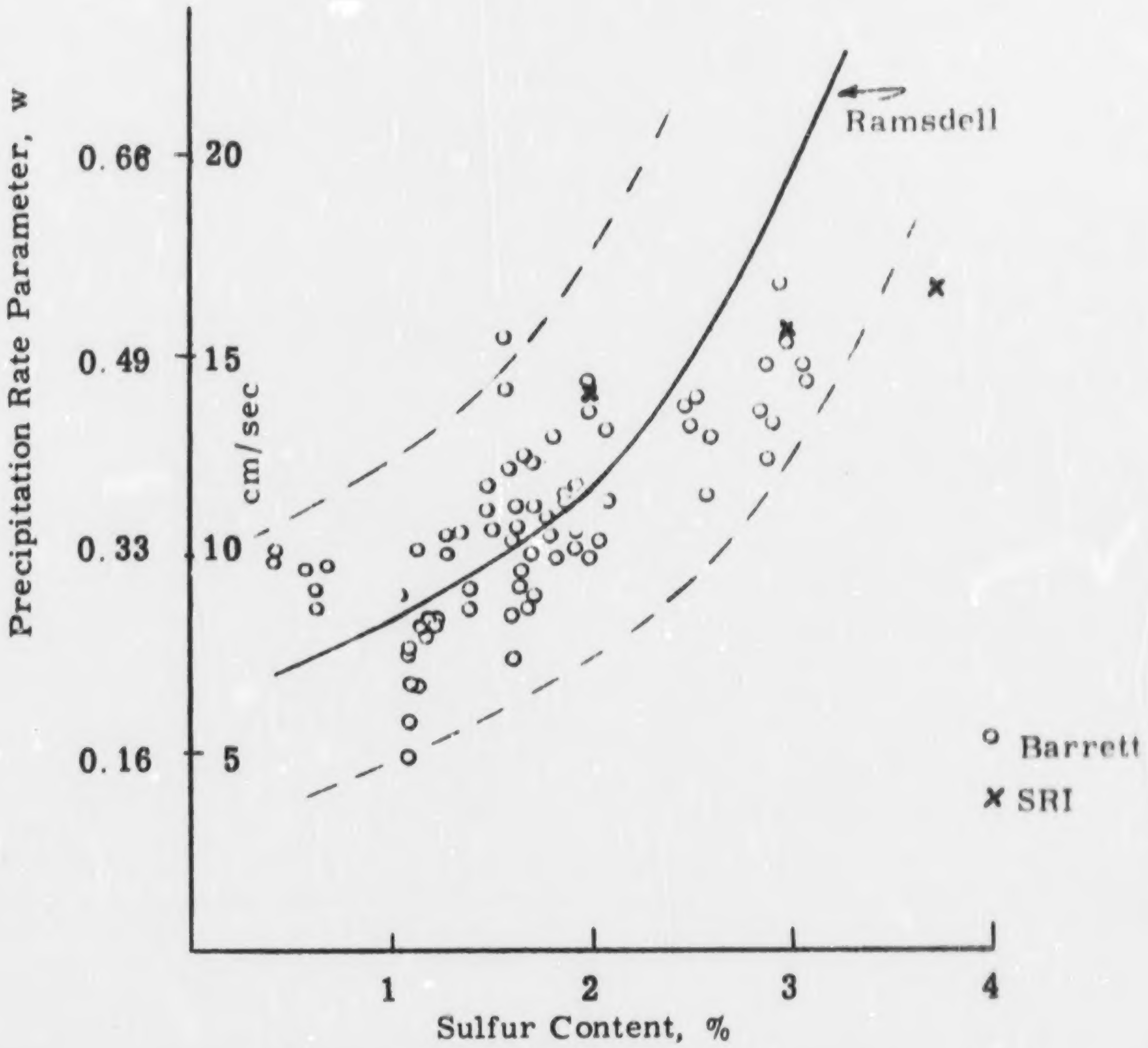


Figure 15.8. Variation in Precipitation Rate Parameter With Sulfur Content of the Coal.

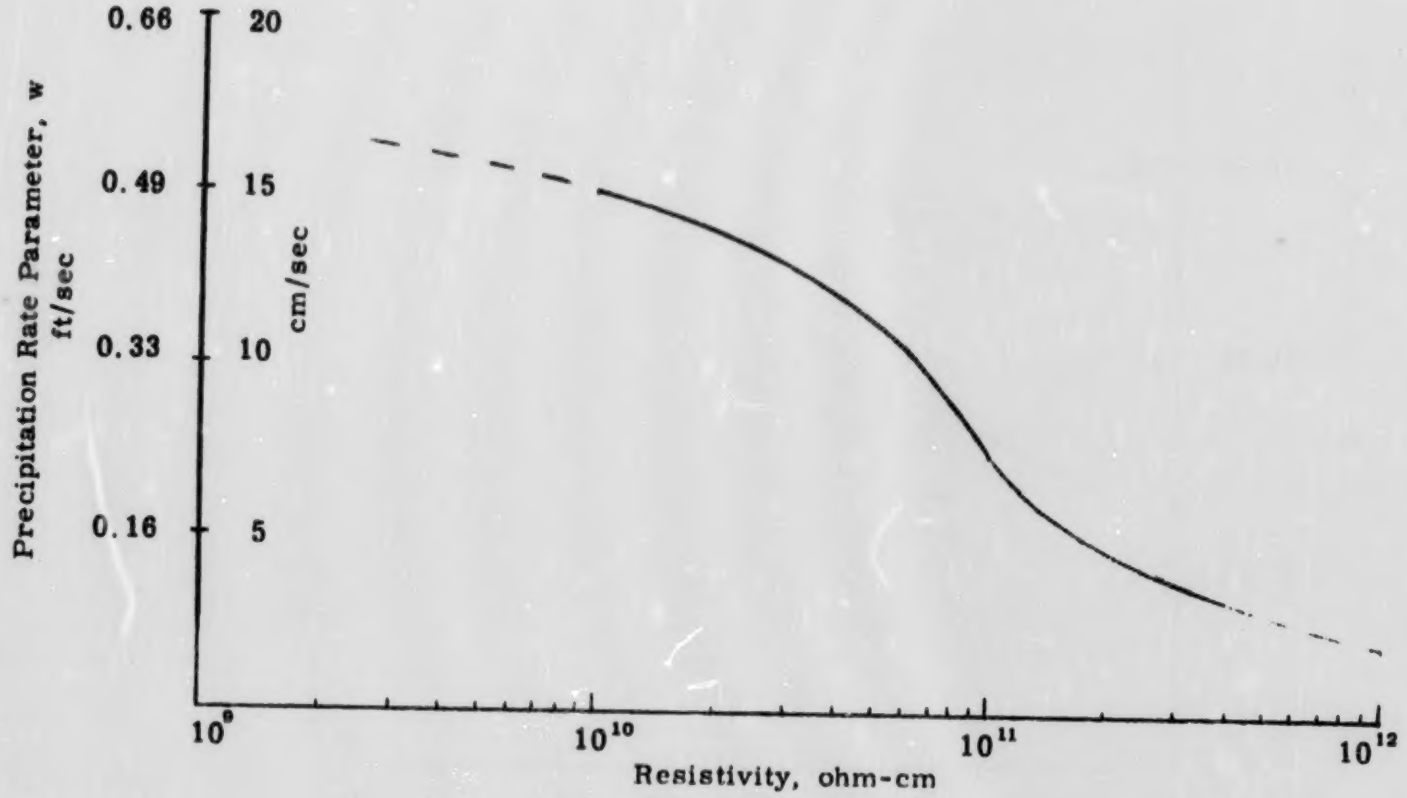


Figure 15.9. Relationship between Precipitation Rate Parameter and In-situ Resistivity.

Designs based on both sulfur content and resistivity are influenced by particle size of the fly ash. Different values of w would apply to precipitators preceded by mechanical collectors than those utilizing precipitators only.

All of the methods of selection of the values of w must be tempered with considerable judgment as related to the particular installation. From a competitive bidding standpoint, the tendency would be to use the highest values of w, in which case the size of the precipitator would be minimal. More conservative design practice would result in selection of a lower value of w, which would result in a larger precipitator with a greater safety margin for error.

Aspect ratio. An important variable is the aspect ratio of the precipitator (height to length of gas passage). Space requirements often determine the overall precipitator dimensions. However, where possible, the aspect ratio should be chosen that will result in ample opportunity for the reentrained dust from the first sections to be recollected. All other factors being equal, lower height to length ratios give better performances.

Precipitator collection plates are made in standardized size ranges, typically 20 - 24 - 30 ft height by 3 - 4 ft length. Once the collection area is selected, the design incorporates collecting plate sections to give the nearest surface area to that calculated.

A survey of a number of installations indicates that plate heights vary from around 17 to 30 feet. Plate lengths associated with 17 foot plate heights range from 12 to 18. With 24 foot high plates, lengths vary from 12 to 27 feet and 30 foot high plates are used with plate length of from 15 to 18. Aspect ratios for these installations varied from 2 to about 9.

Electrical Energization. Selection of the size and type of electrical energization equipment also follows several procedures. The current densities or corona current per length of discharge electrode can be selected by analogy to existing plants, or selection can be based on cumulative experience. Figure 15.10 relates efficiency to corona power per 1,000 cfm for a group of fly ash precipitators. The data apply to delivered power to the discharge electrodes and represent experimental data

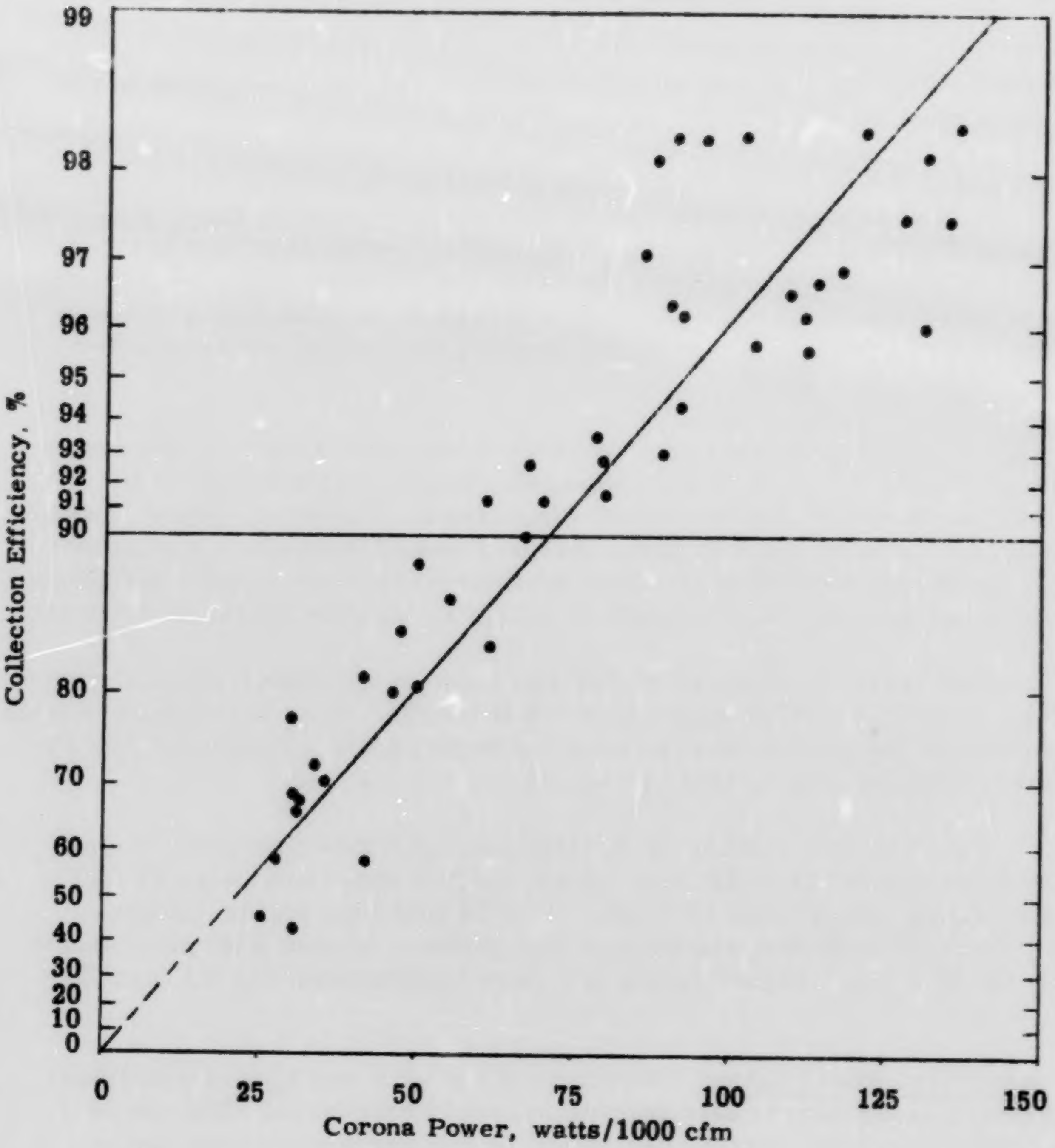


Figure 15.10. Relationship between Collection Efficiency and Corona Power for Fly Ash Precipitators (Test Results).

taken in the field on full size precipitators. This curve is useful in assessing the adequacy of the corona power for a given installation. Power supplies are generally around 60% efficient so that input power would be approximately 140% of the delivered corona power.

Figure 15.11 is a plot of precipitation rate parameter vs. power per 1,000 sq ft of collecting surface. These two curves can be used to arrive at the total collecting surface area and the power requirements. The collecting surface should agree within a reasonable error with that computed by the Deutsch-Anderson equation for comparable values of w.

Figure 15.12 is a plot of the number of independent bus sections per 1,000 cfm vs. efficiency for a group of fly ash precipitators. The curve is taken from data presented by Ramsdell and the points represent data from various other installations.

The importance of sectionalization of the power supply has been pointed out in the chapter on design methodology. It is of greatest significance in the case of fly ash precipitators because of the large gas flows and large precipitator plants. Sectionalization of the power is important from several standpoints. First, if the precipitator is operating in a sparking mode, increased sectionalization will cause less of the precipitator to be disabled during the interval of the spark. This results in higher average voltage, higher electric field, and better precipitation. Also, the smaller electrical sets have higher internal impedances which give better spark quenching and minimize the tendency of a spark to develop into an arc. Third, effects of localized electrodes misalignments are limited to smaller precipitator sections thereby permitting higher voltages in the remaining sections. Finally, in very large precipitators, reasonably good collection efficiencies can still be maintained even if one section has to be deenergized because of wire breakage, or other electrical trouble.

Increasing the number of electrical sections leads to increased costs for an installation because the cost of the high voltage power supply is not linearly related to power handling capability. The greater portion of the cost is in providing the high voltage equipment. Increased power can then be provided by using larger components. Hence it is less expensive to provide fewer large power supplies than to power the precipitator from more small sets. However, because of the lower average

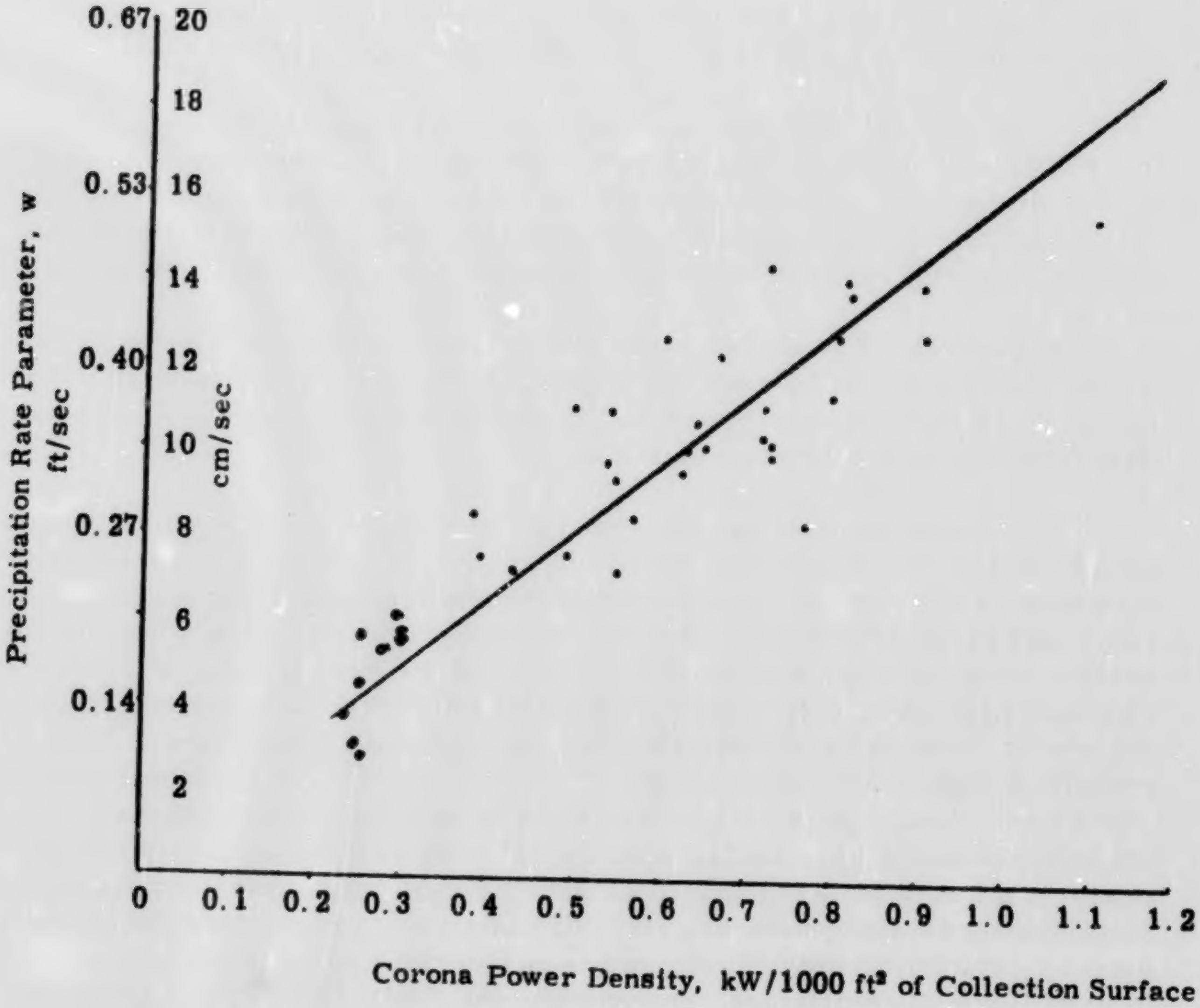


Figure 15.11. Relationship between Precipitation Rate Parameters and Corona Power Density in Electric Utility Installations (Performance Data).

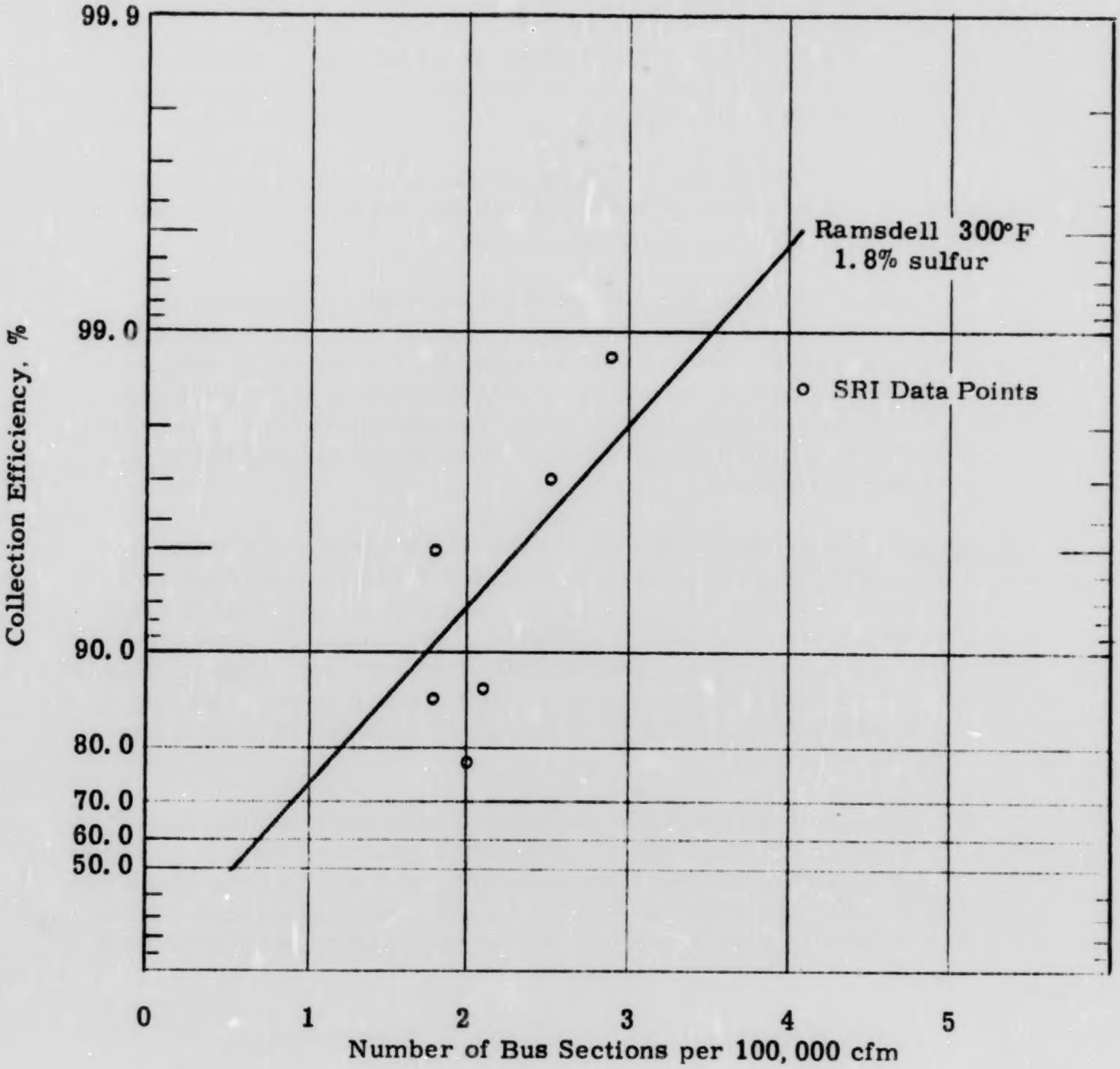


Figure 15.12. Variation in Efficiency with Degree of Sectionalization.

voltage, the precipitation rate parameter would be lower, and the necessity for providing larger collecting surface area would partially offset the lower cost of the larger set.

Power supplies are built in discrete sizes, and when actual requirements of the precipitator are determined, the selection is made based on the nearest integral size available.

Voltages for fly ash precipitators are normally based on plate spacing. Figure 15.13 shows the range of voltage gradients based on the average voltage divided by plate spacing for a group of fly ash precipitators. Plate spacing is chosen so that tolerances in erection and movement of the electrodes make a small percentage change in the electric field. It is a common practice to provide wider spacing for higher electrodes for this reason.

Rappers. The number and size of rappers required for a particular installation vary with precipitator manufacturer and nature of the dust. Variations in the number of sq ft of collecting surface per rapper range from a minimum of around 1200 to 6000 sq ft per rapper. Discharge electrode rappers range from 1000 to 7000 ft of wire per rapper. Size of the rappers varies from around 25 to 50 ft lbs per cycle. Rapping times are adjustable over a range of around 30 to around 600 seconds between raps.

The important consideration in rapping is to provide ample acceleration to dislodge the dust without excessive reentrainment. In general, it has been shown that accelerations measured on the collection electrode of 30 to 50 g's per rap are required for fly ash. Both cycle and rapping intensity are usually adjusted in the field to optimize rapping conditions for maximum precipitator performance.

Effects of gas temperature on fly ash precipitator design.

Low temperature fly ash precipitators

The majority of fly ash precipitators operate at gas temperatures in the neighborhood of 300° F. Referring to Figure 15.4, it can be seen that this temperature is close to the value that would result in maximum resistivity for a 0.5% sulfur coal. Consequently, power plants burning

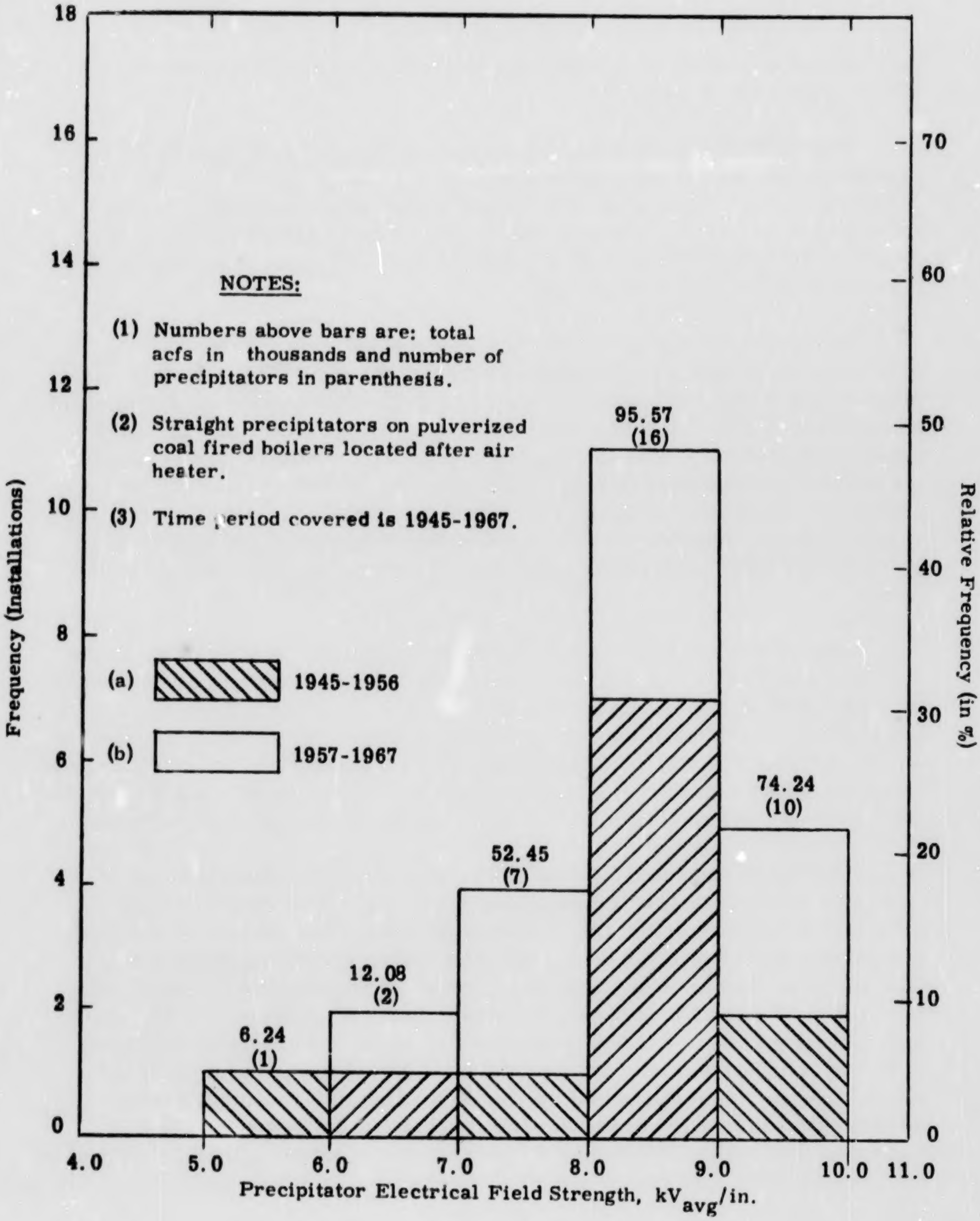


Figure 15.13. Distribution of Design Precipitator Electrical Field Strength.

high ash, low sulfur coal, often experience precipitator problems associated with high resistivity.

One alternative method of improving the performance of fly ash precipitators collecting high resistivity dust is to reduce the gas temperature so that the resistivity is in a range more favorable for precipitation. Figure 15.4 indicates that a reduction in temperature of the order of 30 to 40°F from a 300°F operating point, would make a substantial reduction in resistivity.

A further advantage to decreasing the gas temperature is the increased economy due to more effective heat recovery. A 30 to 40°F drop in gas temperature could result in increased boiler efficiencies of as much as 1.5%. In large power plants, the resultant saving in fuel costs is significant, and the expected tendency would be to operate at the lowest gas temperature possible whether or not resistivity was a problem. However, most power plants burning coal with sulfur in the range of 2-3%, operate at gas temperatures from the air heater in the vicinity of 300°F, primarily to minimize corrosion and fouling tendencies.

The criterion used in selecting the operating temperature is the average value of the exit gas temperature and the entering air temperature at the cold end of the air heater. The lowest permissible cold end temperature is determined by the sulfur content of the coal, and can be as low as 150°F for 0.5% sulfur coal. The entering air temperature can be modified by steam coils, or combustion air can bypass the air heater, thereby reducing the volume of cold air flowing through the air heater.

Corrosion and fouling are associated with the condensation of sulfuric acid on the low temperature surfaces. The rate of condensation, and hence corrosion, depends on the sulfur content of the fuel, moisture content of the flue gases, and boiler operating practice. Clark[12] reported on a study of the corrosion and fouling potential of flue gases of over 60 different boilers. In these studies, corrosion and fouling potential curves were developed for a variety of flue gas compositions for various temperatures. Metal temperatures of the air preheater, superimposed on the corrosion and fouling potential curves, indicate that corrosion and fouling can be restricted to the extreme cold end. By the use of corrosion-resistant alloys or enameled cold end layers, the corrosion problem can be handled, permitting exit gas temperatures as low as 200° F. Figure 15.14 shows the minimum cold end temperature for various sulfur contents and various types of material in the final section of the air preheater.

Surveys of operating practices in several power plants confirm that gas temperatures in the range of 250-260° F are used even with sulfur contents of 2-3%. Examination of the interior of precipitators operating on a few plants shows no evidence of corrosion.

Difficulties have been encountered with the collection efficiencies of precipitators operating at low temperatures on fly ash from high sulfur coal. The exact nature of the problems has not been resolved, but improvements in efficiency have been achieved by increasing the temperature to 300-320° F, injection of ammonia, or use of lower sulfur coal. It has been suggested that the problem with operating with low temperature and high sulfur is a low resistivity fly ash, resulting in excessive reentrainment. Although the problem could be resolved by operation at higher gas temperatures, the reduction of boiler efficiency makes this option unacceptable.

There is considerable variation in operating policy regarding the minimum gas temperature, depending perhaps on local conditions and experience. It would appear, however, that operating at low temperatures to overcome the problem of high resistivity would offer good possibilities, providing control of corrosion and fouling can be maintained.

High temperature fly ash precipitators

The majority of fly ash precipitators are located downstream from the air preheater, and operate at temperatures ranging from around 270 to 320° F. An alternate location of the fly ash removal equipment is ahead of the air preheater. The temperature of the flue gas at this point is in the neighborhood of 700° F.

There are several advantages to locating the precipitator ahead of the air heater. Because of the high temperature, the resistivity of the fly ash will be less than the critical 2×10^{10} ohm-cm regardless of the sulfur or moisture content of the fuel. Consequently, many of the problems

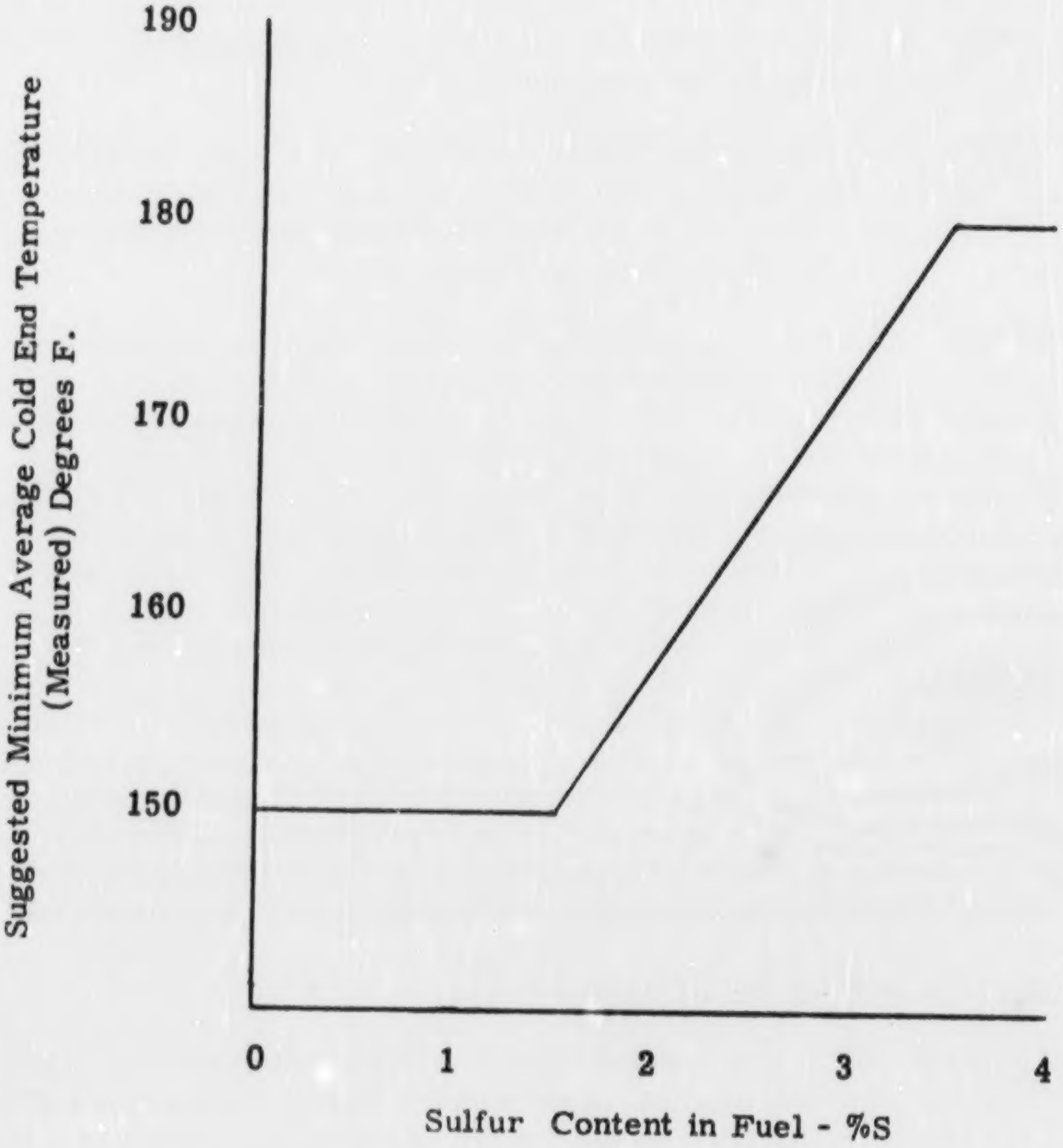


Figure 15.14. Cold End Temperature and Material Selection Guide (after Clark[11]).

associated with the collection of high resistivity dusts will disappear. Also, removal of the fly ash ahead of air preheater minimizes the problems of fly ash fouling. A third advantage is that for units with both oil and coal firing, the problem of collecting and conveying of the oil ash is minimized.

The following factors must be considered in locating the precipitator ahead of the air heater: (1) The gas volume to be handled is increased over that required for operation following the air heater by the ratio of the absolute temperatures. The gas volume at a temperature of 700° F is approximately 1.55 times that at 290° F. (2) The electrical characteristics of the precipitator are modified by higher temperature operation since the lower gas densities result in lower corona voltages and electric fields. (3) Referring to the Deutsch-Anderson equation, the collecting surface area required to achieve a given efficiency would be 1.55 times greater for operation at 700° F than at 290° F assuming the same precipitation rate parameter. (4) The viscosity of the gas increases with temperature, thus lowering w. (5) In the case of high resistivity dust, the precipitation rate parameter can be quite low and in such instances, increased temperature operation can result in a higher precipitation rate parameter that may more than offset the increased size due to increased volume. (6) Under some conditions the fly ash may sinter, producing a difficult to remove deposit.

The only fly ash precipitator operating ahead of the air heater is the Ravenswood Plant of the Consolidated Edison Company. Ramsdell[13] describes this system as a combination electrostatic precipitator followed by a mechanical cyclone collector. The electrostatic collection system consists of four individual precipitators, each composed of four units containing eight electrical sections each. The electrical energization is supplied by sixty-four 70 kVp, 750 mA electrical sets with half-wave silicon diode rectifiers tanked with transformers.

There are 128 individual bus sections, 320 plate rappers, and 128 wire vibrators.

The fuel is a pulverized bituminous coal with 0.6 to 3% sulfur. The precipitator is designed to give 99.2% collection efficiency with 4,300,000 acfm gas flow, 99.8% at 3,230,000 acfm, and 99.9% at 2,150,000 acfm.

The total collecting plate area is 1, 008, 000 sq ft, giving ~~234~~ 430 ft^2/1000 cfm at noted conditions. On this basis the design precipitation rate parameter would be

$$w = \frac{V}{A} \ln \frac{100}{100 - 99.2}$$

$$= 20.6 \text{ ft/min or } 10.5 \text{ cm/sec}$$

Costs of the precipitator installation at the Ravenswood Plant were about \$4.7 million including erection. The complete installation cost was about \$10-12 million including foundation, flues, ash handling, etc. The erected cost per acfm for the precipitator is \$1.09. The volume of gas handled at 300° F would be $V = 4,500,000 \times \frac{300 + 454}{700 + 454} = 2,800,000$ and referring to Figure 15.27, the corresponding average cost would be about \$1.6 million. The cost per acfm would be about \$0.57. This would result in an erected cost about 2.7 times higher for the high temperature installation.

Since this particular plant is the first of its kind, the costs may not be representative of those for subsequent installations.

15.5 SUMMARY OF PRECIPITATOR OPERATING CONDITIONS

Operating conditions for a group of fly ash precipitators installed from 1945 to 1967 have been summarized in a series of histograms given in Figures 15.15 to 15.18. Figures 15.15 and 15.16 show the range of precipitator input parameters including dust loading and gas temperature. Figures 15.17 and 15.18 show the precipitator design parameters including power input, gas velocity, and electric field strength (voltage divided by interelectrode spacing).

Regression analysis. Several attempts have been made to examine the performance of fly ash precipitators in an effort to arrive at a rational basis for evaluating designs and assessing critical factors in performance. Since the number of variables involved in a given installation is large, and since these change from plant to plant, it is impractical to obtain

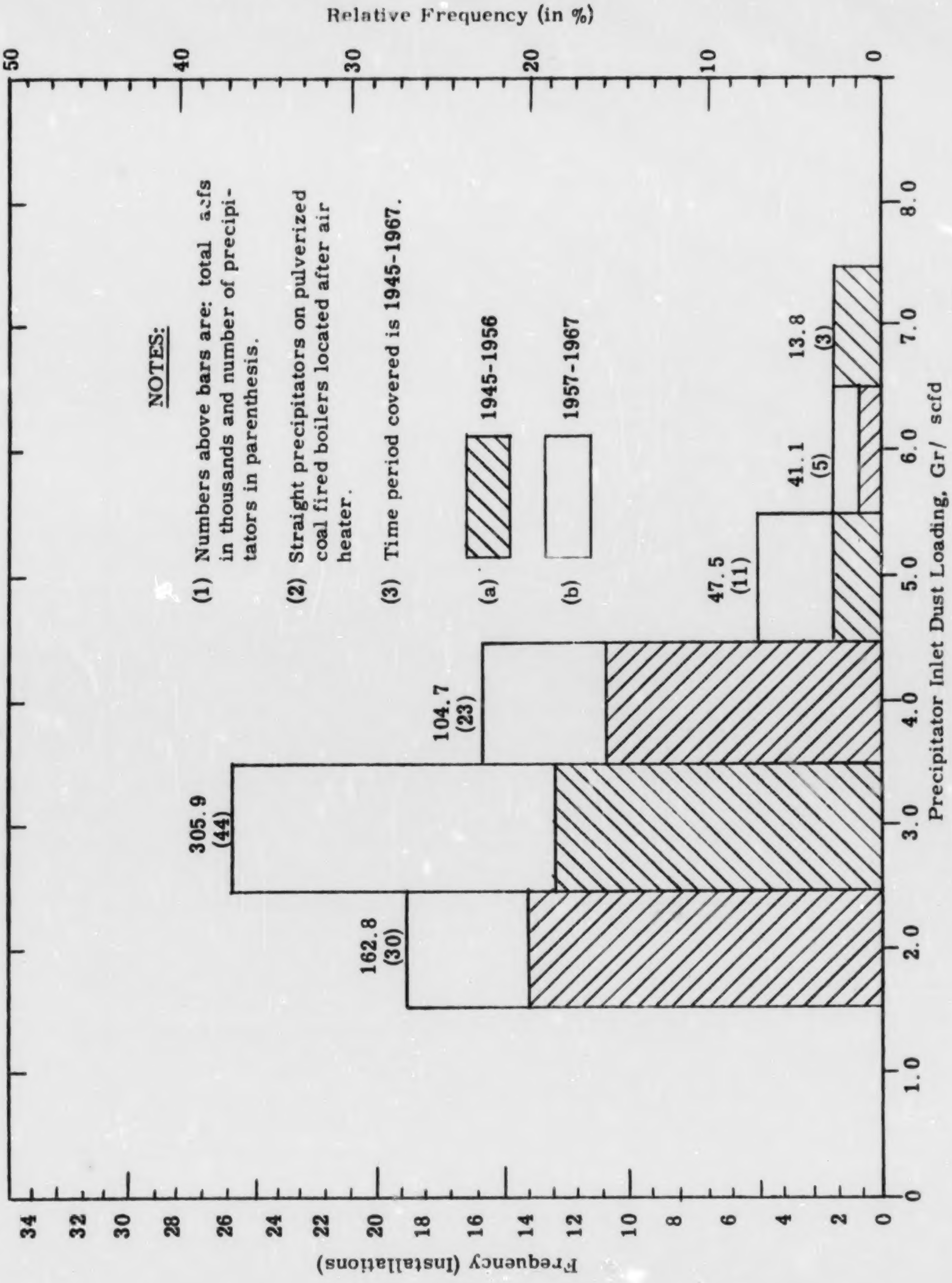


Figure 15.15. Distribution of Precipitator Inlet Dust Loading.

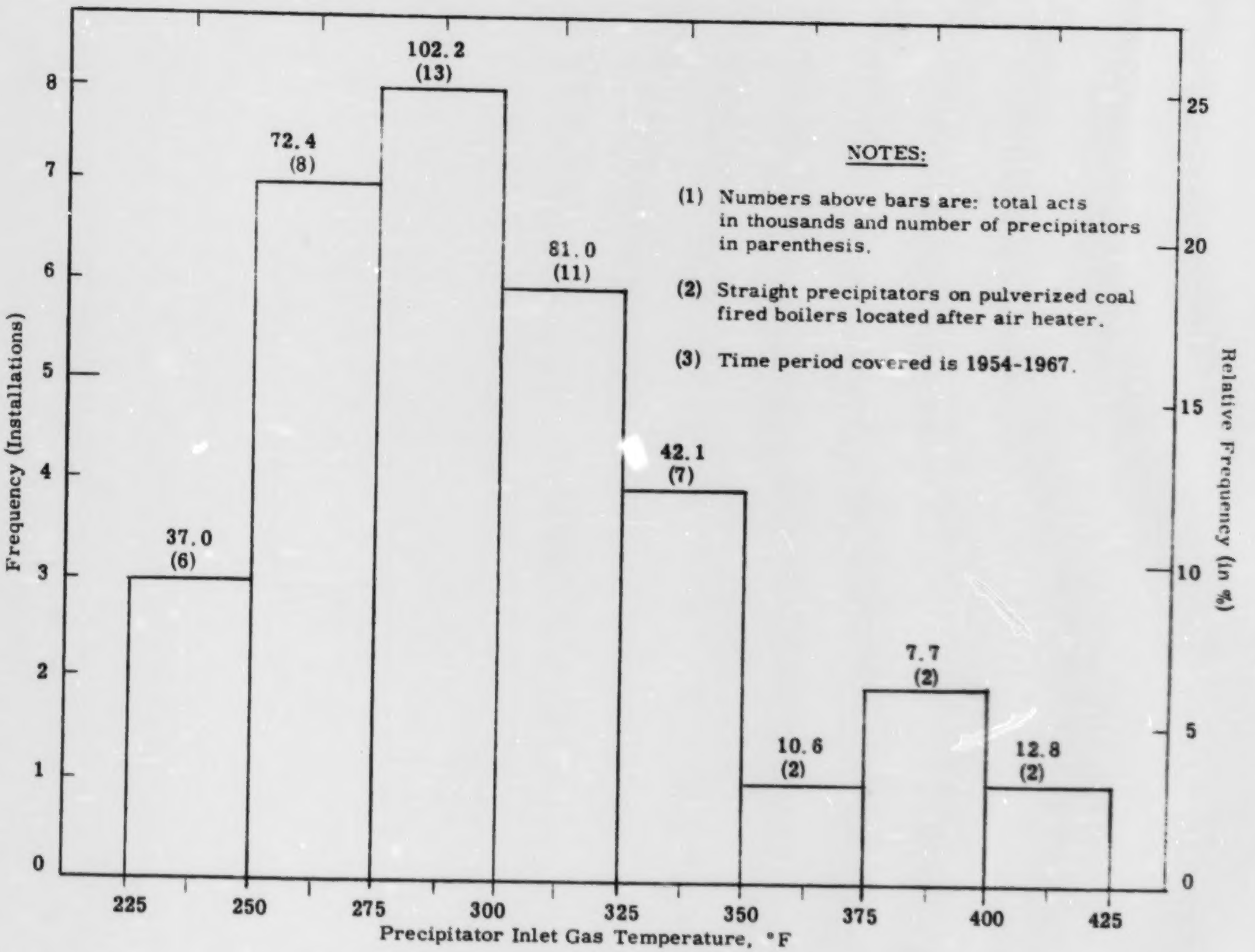
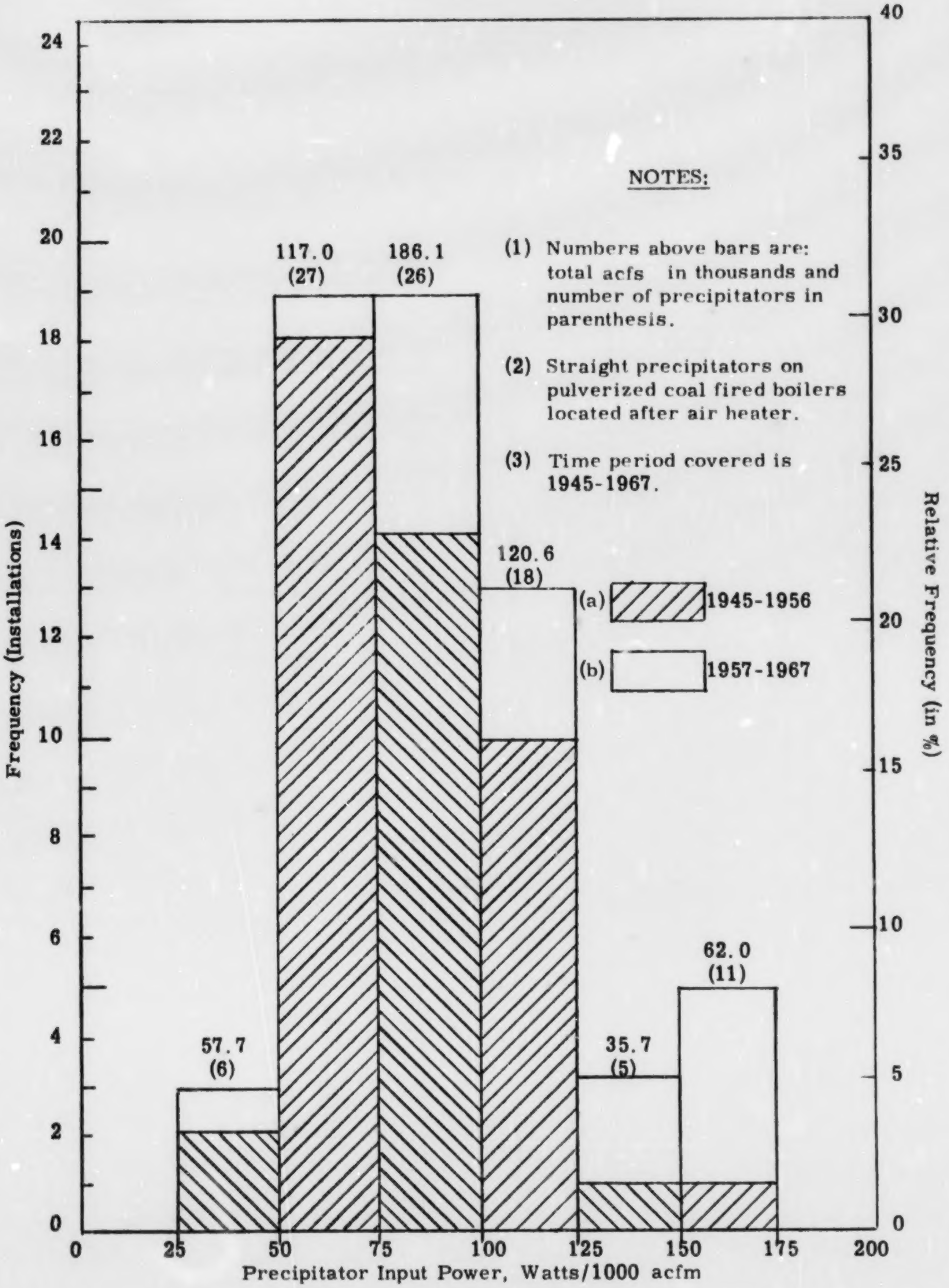


Figure 15.16. Distribution of Precipitator Inlet Gas Temperature.



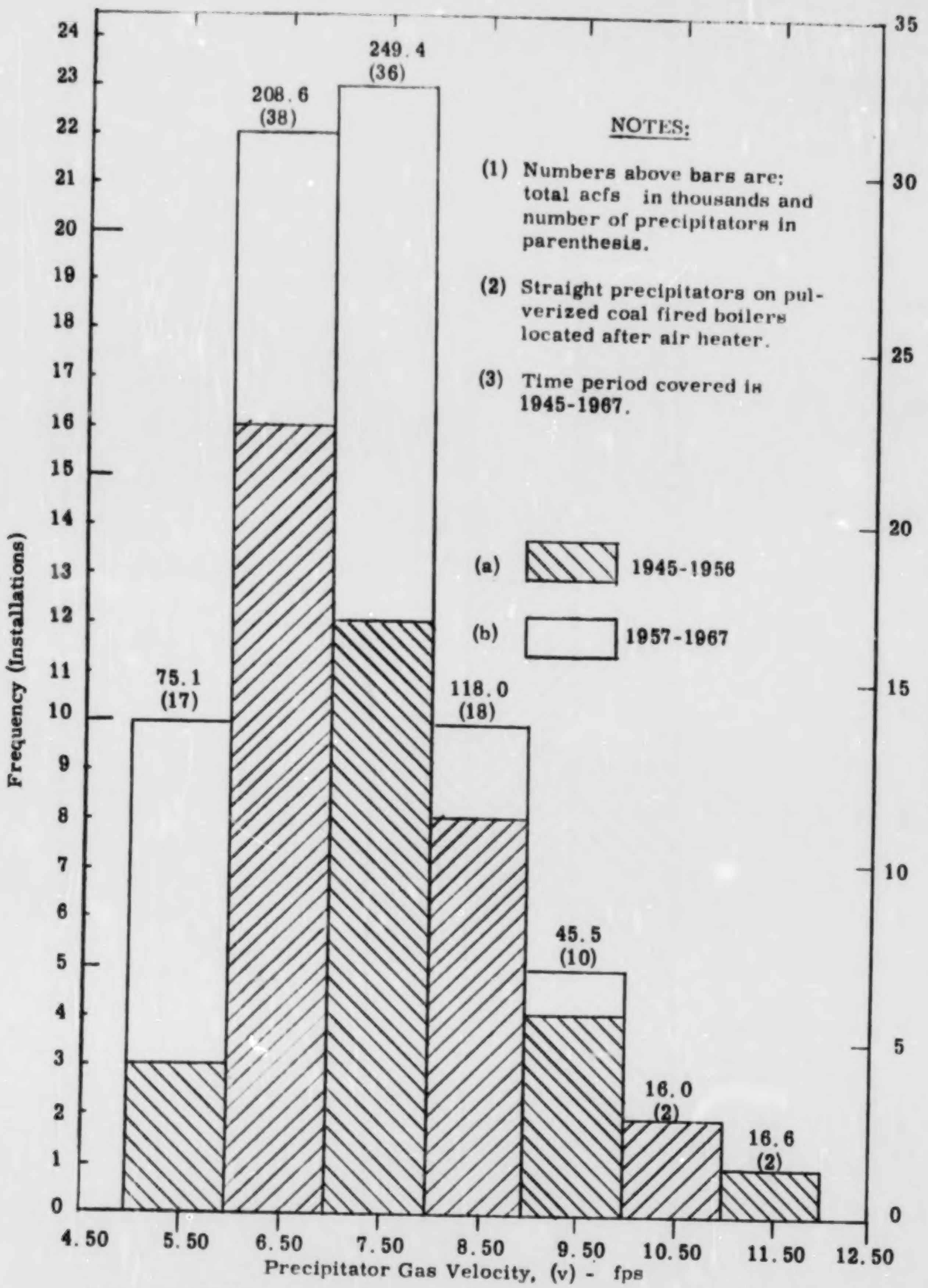
NOTES:

(1) Numbers above bars are: total acfs in thousands and number of precipitators in parenthesis.

(2) Straight precipitators on pulverized coal fired boilers located after air heater.

(3) Time period covered is 1945-1967.

Figure 15.17. Distribution of Design Precipitator Input Power.



NOTES:

(1) Numbers above bars are: total acfs in thousands and number of precipitators in parenthesis.

(2) Straight precipitators on pulverized coal fired boilers located after air heater.

(3) Time period covered is 1945-1967.

Figure 15.18. Distribution of Precipitator Gas Velocity.

sufficient data on the effect of a single variable while maintaining the others constant. Instead, methods for determining the influence of a single variable in a condition in which all variables are changing have to be used. The technique of multiple linear regression analysis has been applied to examine the data from a group of installations and to compare the calculated and measured precipitation rate parameters.

The technique of regression analysis assumes the parameters to be related according to a predetermined equation such as:

$$w = A + A_1 f(d_1) + A_2 f(d_2) \ldots$$

where

A = linear coefficients
$f(d_1)$ = function of various precipitator parameters

To obtain the relationship between the precipitation rate parameter w and the precipitator, a listing of the factors that can influence w is made along with the probable forms of these variables. Factors such as sulfur content, ash content, temperature, etc. are assumed to be important and are entered into a standard regression analysis computer program. The data from all of the installations are then entered and the statistical importance of each term in the equation is tested. Terms that appear to be random are rejected and the coefficients for the remaining terms are computed.

Barrett[14] performed a linear regression analysis of a group of approximately 74 tests conducted on 19 different power plant stations in the United Kingdom and arrived at the following equation:

$$w = 31.2 - 4.8 \ln F - 0.0008P - 7 \times 10^{-8} - 8.1 \exp(-0.5S) - \frac{43.3}{A}$$

where

w = precipitation rate parameter (cm/sec)
P = collection electrode area per unit of high tension capacity (ft^2/kVA)
F = collecting electrode area per unit of gas volume ($ft^2/ft^3/sec$)
g = specific surface area of inlet dust (cm^2/g)
S = total sulfur content of coal as received (%)
A = ash content of coal as received (%)

The data showing the calculated and measured precipitation rate parameters are given in Figure 15.19. The correlation coefficient for the regression analysis is about 0.86.

Using a different set of performance parameters, a regression analysis performed on a group of 49 tests on precipitators for which reasonably accurate test data were available gave the following equation:

$$\begin{aligned} w &= -12 + 0.5j - 5.6P + 0.58V + 0.36N + 0.11M + 0.7 \exp(0.5S) \\ &= -3.5 \times 10^{-3}T - 8 \times 10^{-8}\, g^2 \end{aligned}$$

when

w = precipitation rate parameter (cm/sec)
j = current density $(\mu a/ft^2)$
P = power density $(watts/ft^2)$
V = applied voltage (kV)
N = number of electrical sections
M = moisture content of flue gas (%)
S = sulfur content of fuel (%)
B = dust burden $(grains/ft^3)$
A = ash content of fuel (%)
T = gas temperature (°F)
g = particle surface area/gm (cm^2/gm)

The results of the regression analysis of these data are plotted in Figure 15.20. This technique seems to be good for predicting the precipitation rate parameter when all the factors required for computation are known. The correlation coefficient for the regression equation for this series of 49 tests is 0.92; which is good for data of this nature.

It is important to note that specific values for many of the variables included in the second regression analysis will not be available for new installations. Specifically, values for current density, power density, and applied voltage will not be known until the installation has been completed. Therefore it is important to consider the range of correction that each of the variables is expected to have for normal variations in the performance variables.

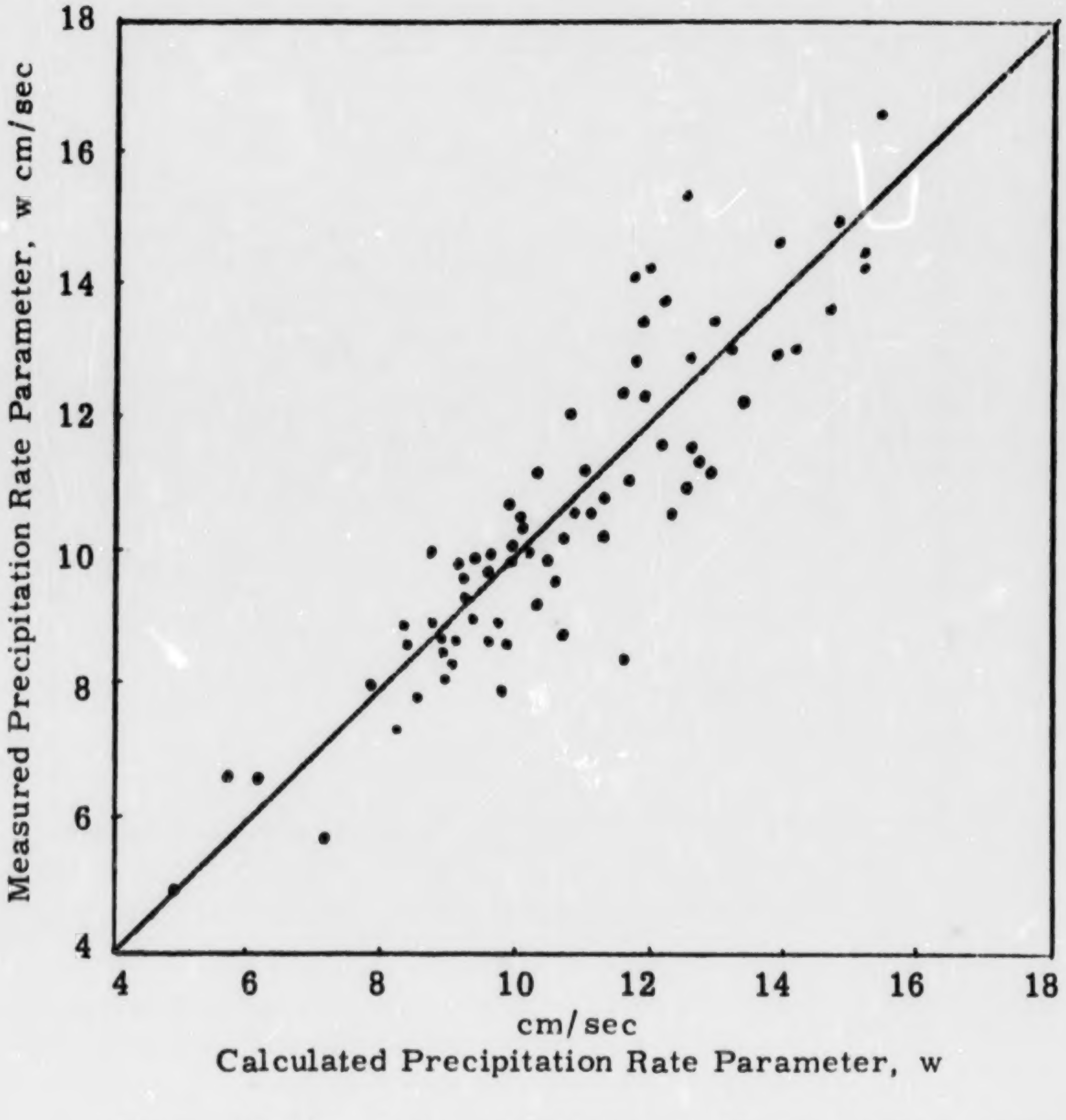


Figure 15.19. Comparison between Measured Precipitation Rate Parameters and Values Calculated by Regression Analysis (from Barrett[13]).

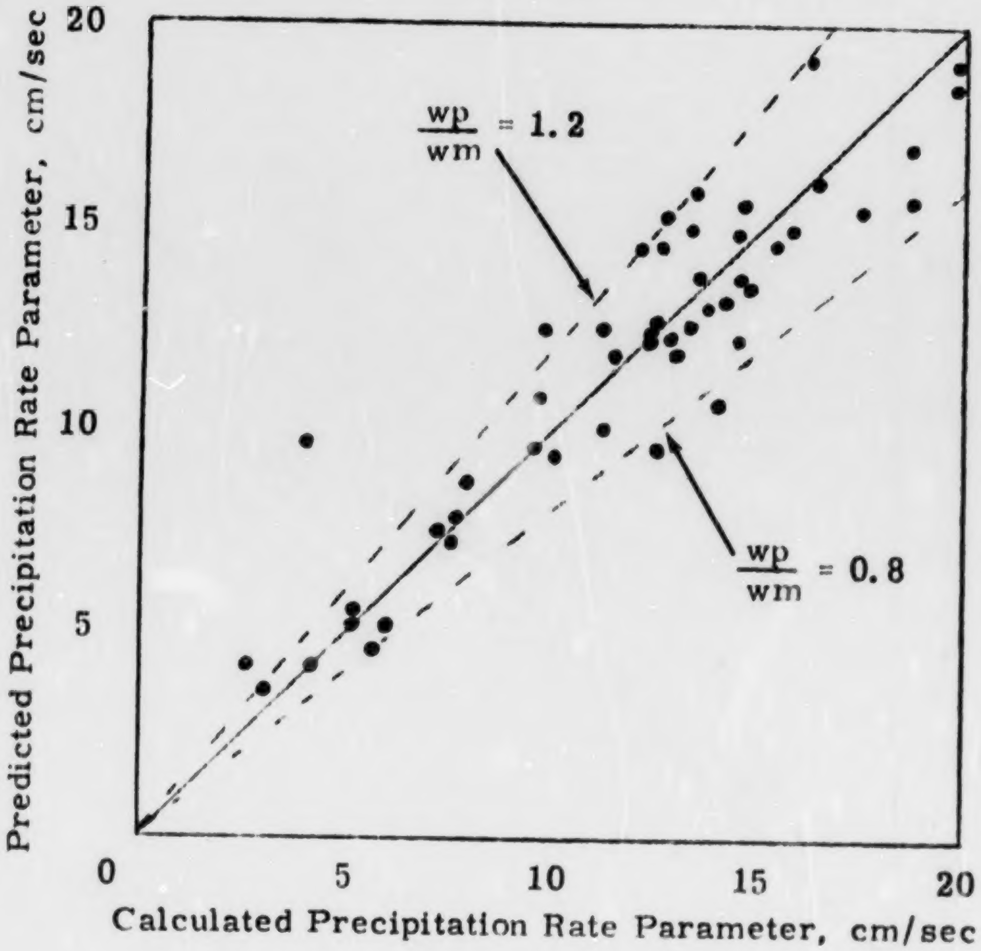


Figure 15.20. Comparison between Measured Precipitation Rate Parameter and Value Calculated from Regression Analysis (SRI).

The current density, applied voltage, and power density are directly related quantities. These three terms must be viewed together. It is to be noted that the product of current density and voltage yields power density. In the equation, the current density and voltage factors increase the precipitation rate parameter, while power density decreases it. The apparent reduction in precipitation rate parameter with increasing power density is surprising until the three power terms are considered together. The current density aids precipitation in two ways:

(1) Increased current density causes reduced charging time.
(2) Increased current density causes increased electric field.

Similarly, the applied voltage influences the electric field directly. Thus, taken individually, current density and applied voltage both increase the precipitation rate parameter.

But it should also be noted that even though the relationships between precipitation rate parameter and the variables need not necessarily be linear, the regression analysis equation will utilize these factors in a linear combination. Thus, the power density term subtracts from the corrections for large values of current and voltage to correct for this nonlinearity in individual effects.

Sectionalization is significant because increased sectionalization leads to increased voltages and currents, which in turn cause an increase in the precipitation rate parameter.

The moisture content, sulfur content, and temperature are significant in that together, they determine, or at least significantly affect, the resistivity of the particulate. As stated previously in this report, the resistivity of the particle is more significant than the above three quantities. But since resistivity values were not known, linear combinations of moisture content, sulfur content, and temperature had to suffice. No method for accurately predicting resistivity from these quantities currently exists.

Higher ash coals can lead to lower precipitation rates because of the effect of higher particle concentrations in increasing the time required to electrically charge the particles.

The surface area per gram of material is a measure of the particle size distribution of the dust with very small particles being characterized by a large value. Since the electrical precipitation rate parameter w is smaller for small particles, dust with large surface area per gram leads to reduced precipitation rate parameters.

The net effect on the precipitation rate parameter for each of the above variables for this particular series of tests is shown in Table 15.4.

15.6 PRECIPITATOR INSTALLATIONS AND ECONOMICS

In order to examine trends in the use of electrostatic precipitation for fly ash collection, estimates of flue gas volumes generated by coal fired power plants were compared with installed precipitator capacities for the period of time from 1923 to 1970. The results are given in Table 15.5 and graphically illustrated in Figure 15.21.

Estimates of flue gas volumes were made utilizing Federal Power Commission coal consumption data and the following assumptions: average carbon in coal - 75%; excess air at the boiler - 25%; air in-leakage - 10%; water vapor volume - 6%; and flue gas temperature at the precipitator - 300° F. The two load factors of 55 and 80% (ratio of average load for one year to the maximum for one hour) bracket assumed load factors for old and new boiler unit. The net result is an estimate of 1.5 and 2.0 cfm of flue gas per ton of coal burned per year for the two load factors. Figure 15.22 is the percentage of flue gas cleaned with electrostatic precipitators based on 80% and 55% load factors.

A two-year delay factor is assumed between the year of contract and year of operation to allow for construction time.

Figure 15.23 shows the installed precipitator gas volume for the years 1923 to 1970 with the annual volume average calculated for five-year periods, while Figure 15.24 illustrates the increasing design efficiency demands for the same time period.

Table 15.4

Variations in Precipitation Rate Parameter Predicted by the Regression Analysis Equation for Fly Ash Precipitators

Variable	Range of Values			Pptn. Rate Parameter Correction, cm/sec		
	Low	Avg	High	Low	Avg	High
Current density ($\mu A/ft^2$)	4.0	17.7	33.6	1.9	8.9	16.8
Power density (w/ft^2)	0.12	0.65	1.56	-0.67	-3.6	-8.7
Voltage (kV)	25.6	35	44	14.8	20.0	25.6
Electrical sets (No.)	3	7.9	18	1.1	2.8	6.5
Moisture content (%)	6	8	15	0.66	0.88	1.65
Sulfur content (%)	1.17	2.43	4.71	0.82	1.7	3.3
Dust burden ($grains/ft^3$)	0.47	2.6	5.94	-0.3	-1.66	-3.8
Ash content (%)	5.8	11.6	14.4	-2.0	-4.0	-5.0
Gas temperature (°F)	248	291	410	-0.9	-1.0	-1.4
Surface area (cm^2/g)	3×10^5	3×10^6	2.7×10^7	-0.02	-0.24	-2.2
Net correction range				+15.4	+23.8	+32.7
Constant term from equation				-12.0	-12.0	-12.0
Range of possible precipitation rate parameter				3.4	11.8	20.7

Table 15.5

Fly Ash Precipitator Installations

Pptr Contract Year	No. of Installations	No. of Pptrs	(1) Total Gas Vol Millions acfm	No. of Installations	Five (5) Yr Periods Total Gas Vol 10^6 acfm	Avg Vol/Yr During Period 10^6 acfm	Pptr Operation Year	Accumul Gas Vol With Pptrs 10^6 acfm	Coal Burned 10^6 tons per year	(2) Total Gas Vol Calculated From Coal Burned 10^6 acfm	Percent of Total Vol with Pptrs
1923	1	3	0.80				1925	0.80	35.6	53.4	1.5
1926	2	3	1.35				1928	2.15	38.0	58	3.7
1927	2	11	3.59	10	8.58	1.72	1929	5.74	41.8	62.8	9.1
1928	1	2	0.22				1930	5.96	40.3	60.5	9.9
1929	4	20	2.62				1931	8.58	36.1	54.2	15.8
1930	6	10	1.43				1932	10.01	28.0	42.0	23.8
1931	2	3	0.46				1933	10.47	28.5	42.8	24.5
1932	1	2	0.37	13	3.46	0.69	1934	10.84	31.4	47.0	23.1
1933	2	4	0.68				1935	11.52	32.7	49.0	23.5
1934	2	4	0.52				1936	12.04	40.1	60.0	20.1
1935	2	5	0.79				1937	12.63	42.9	64.5	19.9
1936	13	23	3.59				1938	16.42	38.4	57.6	28.5
1937	15	29	4.71	39	12.68	2.54	1939	21.13	44.5	66.8	31.6
1938	1	1	0.25				1940	21.38	51.5	77.3	27.7
1939	8	17	3.34				1941	24.72	62.6	94.0	26.3
1940	22	36	5.42				1942	30.14	66.2	99.3	30.4
1941	31	52	7.84				1943	37.98	77.3	116.0	32.7
1942	9	29	3.37	71	20.28	4.06	1944	41.35	80.1	120.0	34.5
1943	1	3	0.96				1945	42.31	74.7	112.0	37.8
1944	8	11	2.69				1946	45.00	72.2	108.3	41.6
1945	13	17	2.78				1947	47.78	89.5	134.0	35.7
1946	22	39	6.12				1948	53.90	99.6	149.5	36.1
1947	28	59	12.67	105	36.87	7.37	1949	66.57	83.9	126.0	52.8
1948	22	40	9.97				1950	76.54	91.8	138.0	55.5
1949	20	25	5.33				1951	81.87	105.7	159.0	51.5
1950	19	35	6.88				1952	88.75	107.1	161.0	55.1
1951	28	61	14.12				1953	102.87	115.9	174.2	59.1
1952	19	25	7.60	98	42.35	8.47	1954	110.47	118.4	178.0	62.1
1953	23	37	9.27				1955	119.74	143.8	216	55.4
1954	9	15	4.48				1956	124.22	158	237.5	52.3
1955	23	36	14.04				1957	138.26	162	244	56.7
1956	22	37	26.57				1958	164.83	-	-	-
1957	26	43	17.40	107	80.97	16.19	1959	182.23	-	-	-
1958	13	19	7.55				1960	189.78	176.2	265	71.6
1959	23	37	5.41				1961	205.19	-	-	-
1960	17	26	11.09				1962	216.28	-	-	-
1961	12	16	7.56				1963	223.84	211	317	70.6
1962	16	28	17.07	76	68.10	13.62	1964	240.91	-	-	-
1963	15	21	12.54				1965	253.45	244.8	368	68.9
1964	16	27	19.84				1966	273.29	266.5	401	68.2
1965	33	49	27.17				1967	300.46	274.0	412	72.9
1966	48	80	57.24				1968	357.70	-	-	-
1967	55	117	72.51	222	257.61	51.52	1969	430.21	-	-	-
1968	46	106	58.76				1970	488.97	-	-	-
1969	40	67	42.93				1971	531.90	-	-	-
Grand Totals	741	1331	531.9	741							

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(1) Includes all fly ash precipitators except rebuilds.
(2) Based on an 80% load factor.

Table 15.5 (continued)

Pptr Operation Year	Accumul Gas Volume With Pptrs 10^6 acfm	Coal Burned 10^6 tons Per Yr	(1) Total Gas Vol - Calc From Coal Burned 10^6 acfm	% of Total Vol With Pptrs	Elec Energy kW-Hr by Coal Billions	Pounds Coal Per kW-Hr	Total Elec Energy kW-Hr Billions	% by Coal
1925	0.80	35.6	71.2	1.1	35.1	2.0	61.5	57.1
1928	2.15	38.0	76.0	2.8	43.9	1.73	82.8	53.1
1929	5.74	41.8	83.6	6.9	50.4	1.66	92.2	54.7
1930	5.96	40.3	80.6	7.4	50.5	1.60	91.1	55.4
1931	8.58	36.1	72.2	11.9	47.7	1.52	87.3	54.6
1932	10.01	28.0	56.0	17.9	37.7	1.49	79.4	47.5
1933	10.47	28.5	57.0	18.7	39.0	1.46	81.7	47.8
1934	10.84	31.4	62.8	17.3	43.4	1.45	87.3	49.7
1935	11.52	32.7	65.4	17.7	45.4	1.44	95.3	47.6
1936	12.04	40.1	80.2	15.0	55.8	1.44	109.3	51.0
1937	12.83	42.9	85.8	14.9	59.7	1.44	118.9	50.2
1938	16.42	38.4	76.8	21.4	54.7	1.40	113.8	48.2
1939	21.13	44.5	89.0	23.6	64.7	1.38	127.6	50.8
1940	21.38	51.5	103.0	20.6	76.8	1.34	141.8	54.2
1941	24.72	62.6	125.2	19.7	93.8	1.34	164.8	57.0
1942	30.14	66.2	132.4	22.8	101.9	1.30	186.0	54.7
1943	37.98	77.3	154.6	24.5	119.2	1.30	217.7	54.9
1944	41.35	80.1	160.2	25.8	124.1	1.29	228.2	54.4
1945	42.31	74.7	149.4	28.3	114.8	1.30	222.5	51.6
1946	45.00	72.2	144.4	31.2	111.7	1.29	223.2	50.1
1947	47.78	89.5	179.0	26.8	137.0	1.31	255.7	53.3
1948	53.90	99.6	199.2	27.0	152.9	1.30	282.7	54.1
1949	66.57	83.9	167.8	39.6	135.5	1.24	291.1	46.5
1950	76.54	91.8	183.6	41.6	154.5	1.19	329.1	47.0
1951	81.87	105.7	211.4	38.6	185.2	1.14	370.7	49.2
1952	88.75	107.1	214.2	41.2	195.4	1.10	399.2	49.0
1953	102.87	115.9	231.8	44.2	218.9	1.06	442.7	49.5
1954	110.47	118.4	236.8	47.0	239.1	0.99	471.7	50.6
1955	119.74	143.8	287.6	41.5	302.0	0.955	553.5	54.6
1956	124.22	158.0	316.0	39.2	340.0	0.935	601.0	56.7
1957	138.26	162.0	324.0	42.5	350.5	0.925	629.0	55.7
1958	164.83	-	-	-	-	0.900	-	-
1959	182.23	-	-	-	-	0.885	-	-
1960	189.78	176.2	352.4	53.6	403.0	0.875	753.4	53.3
1961	205.19	-	-	-	-	0.870	-	-
1962	216.28	-	-	-	-	0.865	-	-
1963	223.84	211.0	422.0	52.8	491.0	0.860	-	-
1964	240.91	-	-	-	-	0.860	-	-
1965	253.45	244.8	489.6	51.6	571.0	0.860	1055.2	54.0
1966	273.29	266.5	533.0	51.6	621.0	0.860	1144.4	54.1
1967	300.46	274.0	548.0	54.6	637.0	0.860	1211.7	52.8
1968	357.70	-	-	-	-	-	-	-
1969	430.21	-	-	-	-	-	-	-
1970	488.97	-	-	-	-	-	-	-
1971	531.90	-	-	-	-	-	-	-

(1) Based on 55% Load Factor.

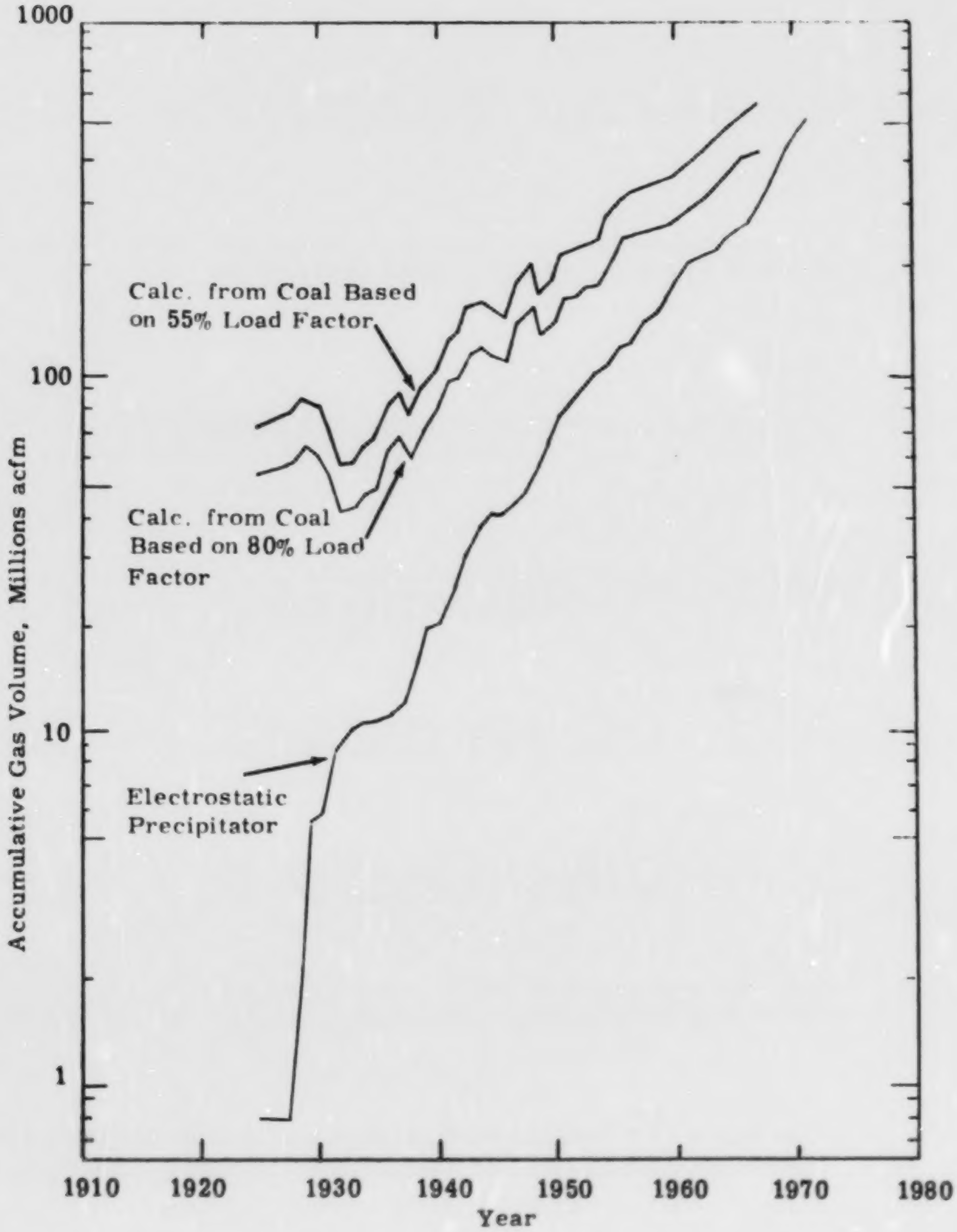

Figure 15.21. Comparison of Gas Volume with Electrostatic Precipitator Cleaning to Total Gas Volume Calculated from Coal Burned.

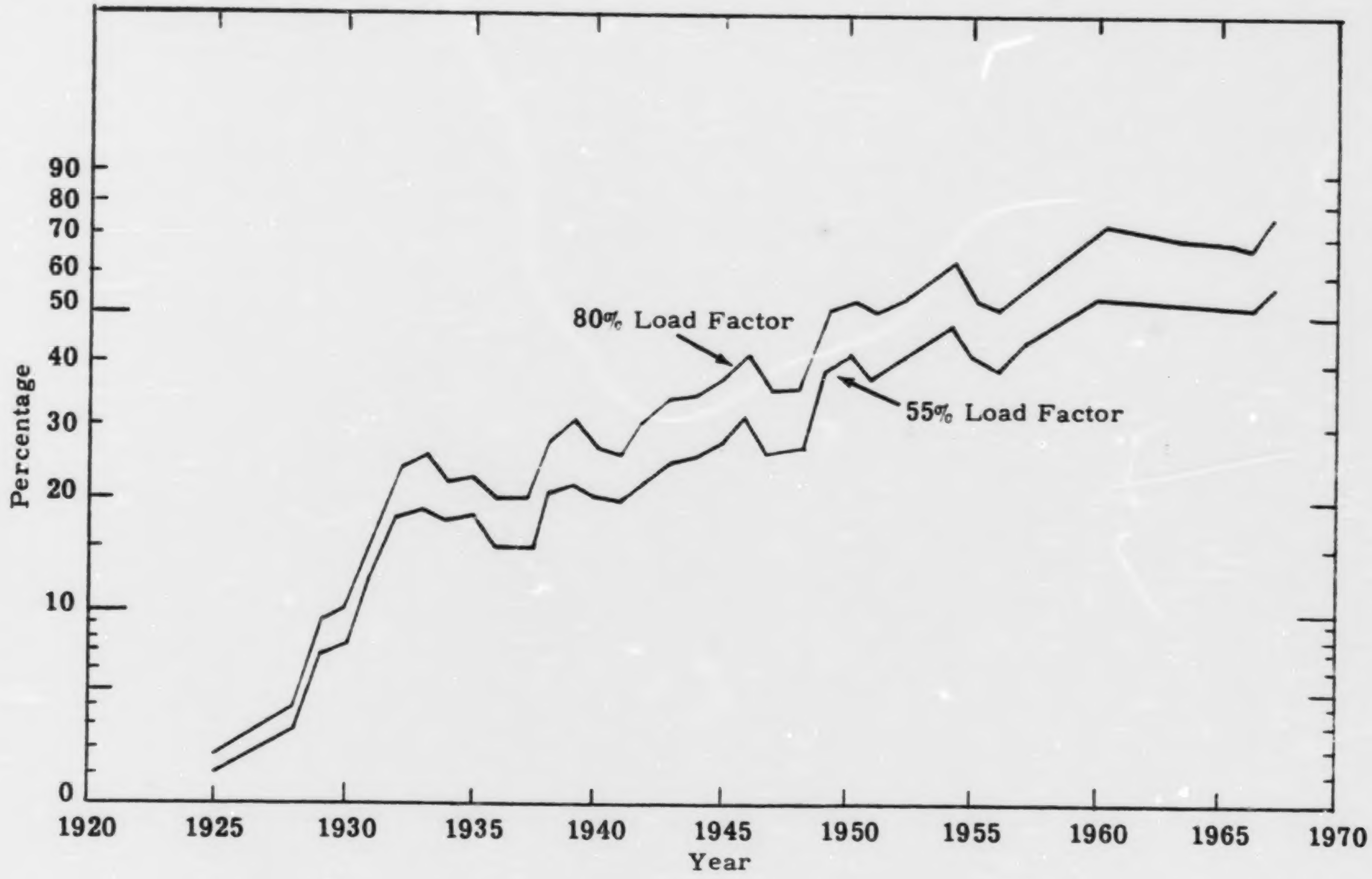

Figure 15. 22. Percentage of Flue Gas from Coal-Firing with Electrostatic Precipitator Cleaning.

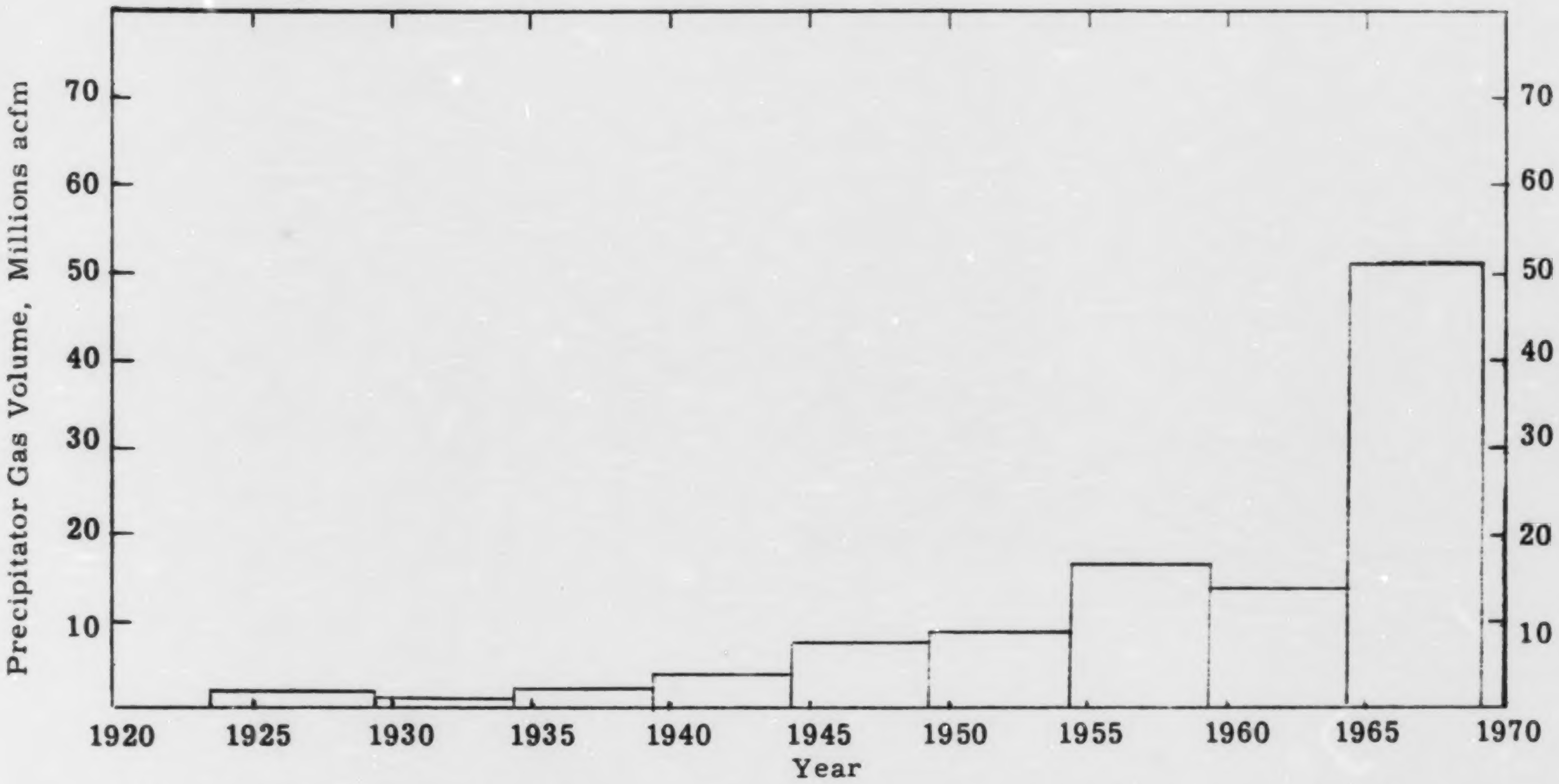

Figure 15.23. Average Yearly Precipitator Gas Volume for the 5-Year Periods from 1923 through 1969.

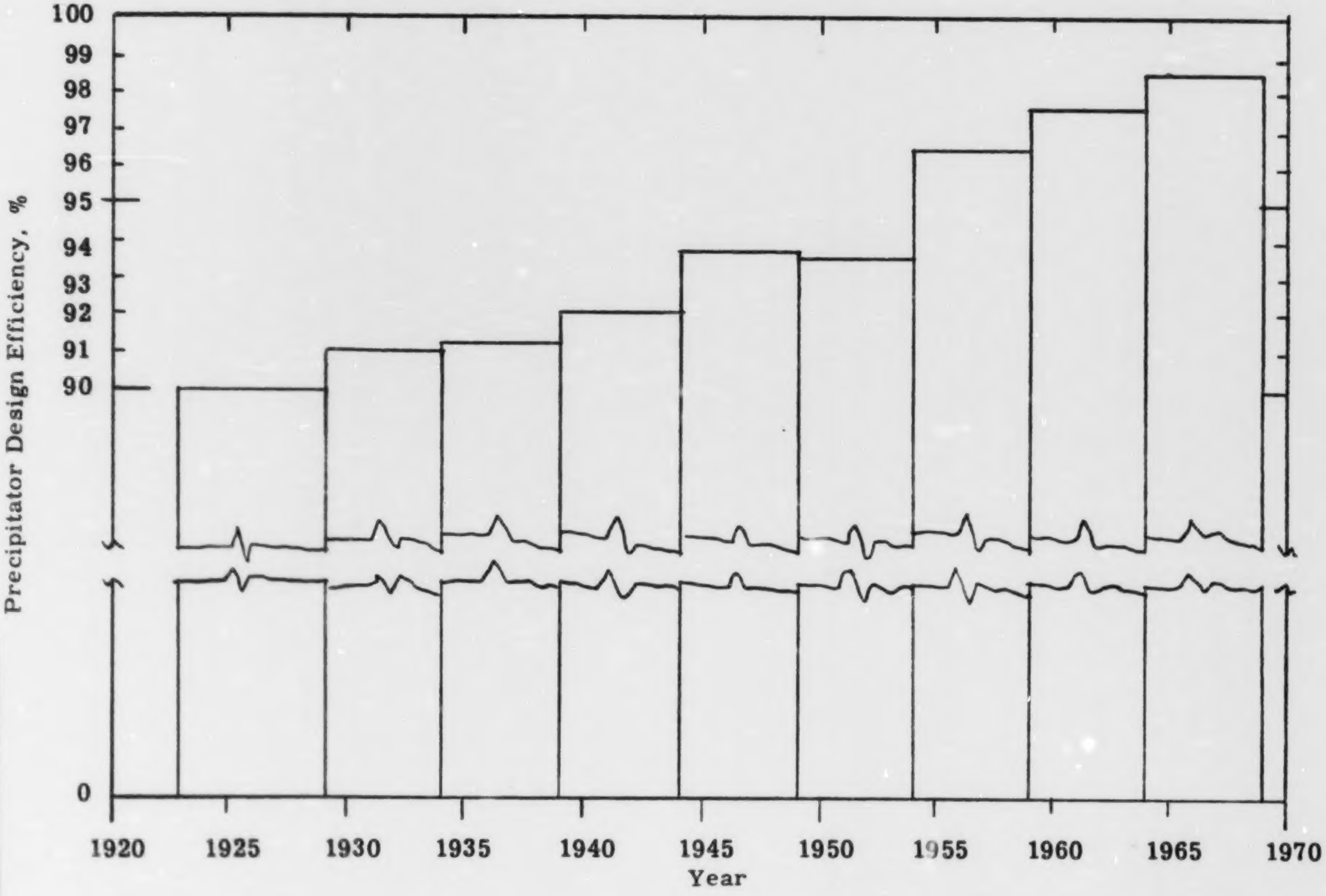

Figure 15. 24. Average Yearly Precipitator Design Efficiency, Percent (Prorated on a cfm Basis) Over the Period 1923 through 1969.

Economics. Fly ash precipitator costs over the past decade typically range from approximately \$0.50 to \$1.00 per cfm of gas treated, on an erected basis. The cost per cfm generally decreases with increasing precipitator gas flow capacity or size, and increases with increasing collection efficiency. Table 15.6 is a tabulation of precipitator costs for various efficiency ranges and for various gas volumes.

Figures 15.25 through 15.27 graphically summarize costs for fly ash precipitator installations for the period of 1959 through 1969. Erected costs include foundation, erection, and electrical work; however, none of the costs include ductwork or air flow regulations other than inlet diffusion plates.

The spread in the cost data at a given efficiency and gas volume can be attributed to many variables, the more important ones being:

(1) The size of the precipitator as related to dust characteristics (resistivity, particle size) and gas conditions (temperature which indirectly affects dust resistivity, moisture content, sulfur oxide content). These variables are reflected in the precipitation rate parameter, which can vary by a factor of 3, and result in a collection surface area that varies over the same range.

(2) The geographical location of the installation, particularly for erected cost since labor costs can vary greatly throughout the country.

(3) The pricing-profit policy of the various competitors bidding on the job; whether the installation is a backfit or upgrading of an existing installation which may require additional engineering costs to "shoehorn" the precipitator into the overall installation; and whether the installation is ground level, above ground level, indoors, outdoors, etc.

Table 15.7 is a summary of the FOB and erected costs of fly ash precipitators from 1960 to 1969. Figure 15.28 shows the average precipitator costs (\$/cfm) for the period 1927-1969. The trend towards higher costs per unit of gas flow reflects the increase in design efficiency as well as increases in production and erection costs.

Table 15.6

Precipitator Costs (1960 through 1964)

Efficiency Range %	Gas Volume Range - Millions acfm							
	0 - 0.249		0.250 - 0.499		0.500 - 0.999		> 1.00	
	F O B	Erected	F O B	Erected	F O B	Erected	F O B	Erected
90.0 - 94.9	111.7 (1)	151.7 (2)	No Data	42.1 (3)	27.9 (2)	54.4 (4)	No Data	No Data
95.0 - 98.9	43.3 (2)	65.2 (1)	35.0 (8)	56.9 (5)	35.6 (14)	47.2 (5)	29.2 (8)	No Data
99.0+	No Data	No Data	No Data	No Data	55.2 (2)	66.8 (2)	33.6 (5)	No Data

Precipitator Costs (1965 through 1969)

Efficiency Range, %	Gas Volume Range - Millions acfm							
	0-0.249		0.250-0.499		0.500-0.999		>1.00	
	F O B	Erected	F O B	Erected	F O B	Erected	F O B	Erected
90.0-94.9	64.7 (2)	59.3 (1)	33.8 (1)	101.1 (6)	25.2 (1)	69.2 (2)	25.8 (3)	No Data
95.0-98.9	75.4 (15)	91.6 (5)	61.1 (10)	88.6 (12)	37.1 (6)	62.4 (9)	31.6 (21)	55.3 (16)
99.0+	98.0 (4)	204.5 (5)	62.7 (4)	103.5 (5)	44.3 (9)	82.7 (9)	34.2 (23)	55.5 (17)

NOTES: (a) Costs are cents/acfm.

(b) Numbers in parentheses are sample size, i.e., number of installation contract prices averaged to obtain costs/acfm (all precipitator manufacturers represented).

(c) Costs are for period shown, no reduction to base year.

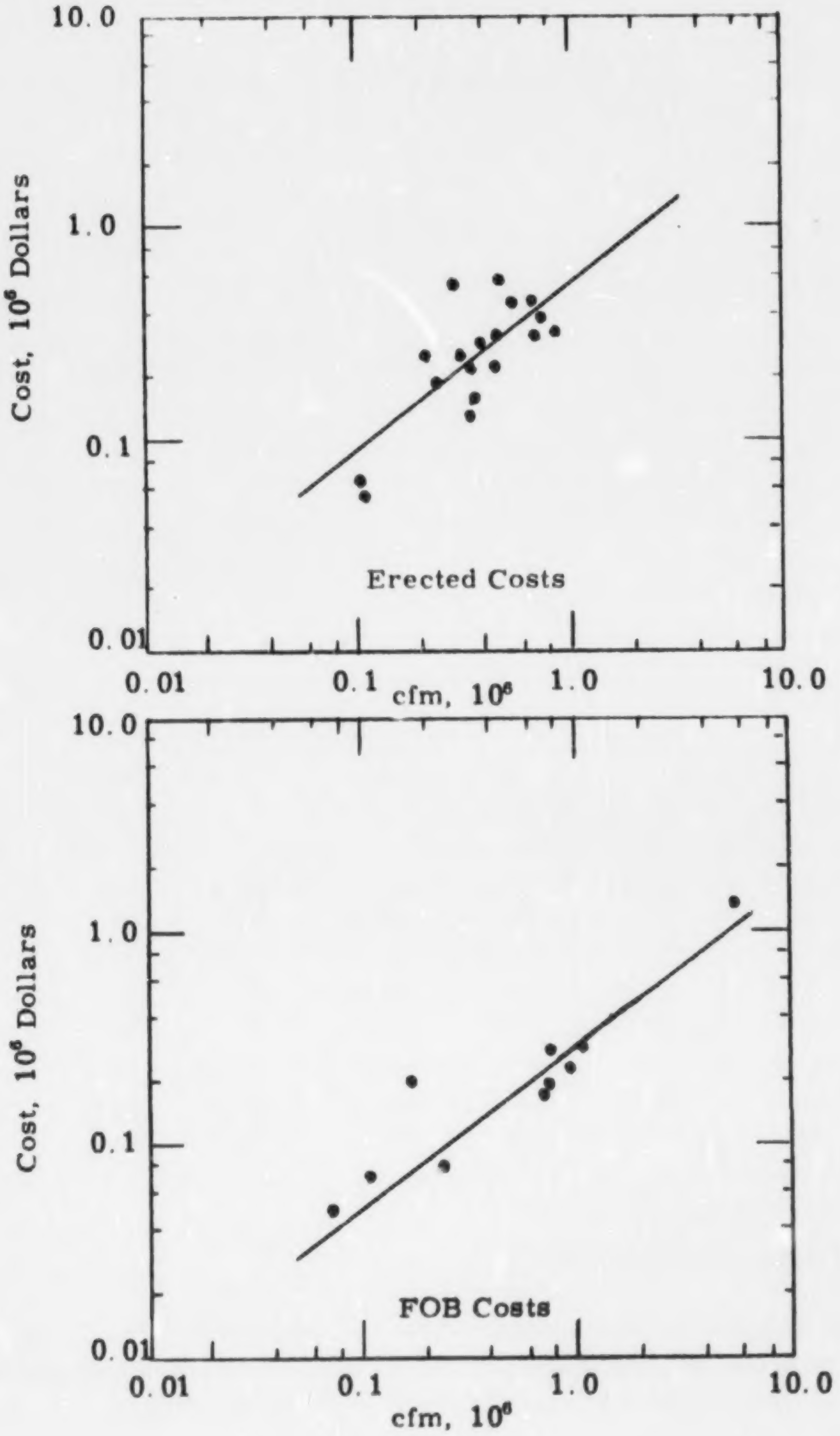

Figure 15.25. Electric Utility Electrostatic Precipitator Costs Vs. Volume Flow for Collection Efficiencies between 90 and 95%. These Costs Represent Installations between 1959 and 1969. The Costs Were Corrected to 1963 Dollars.

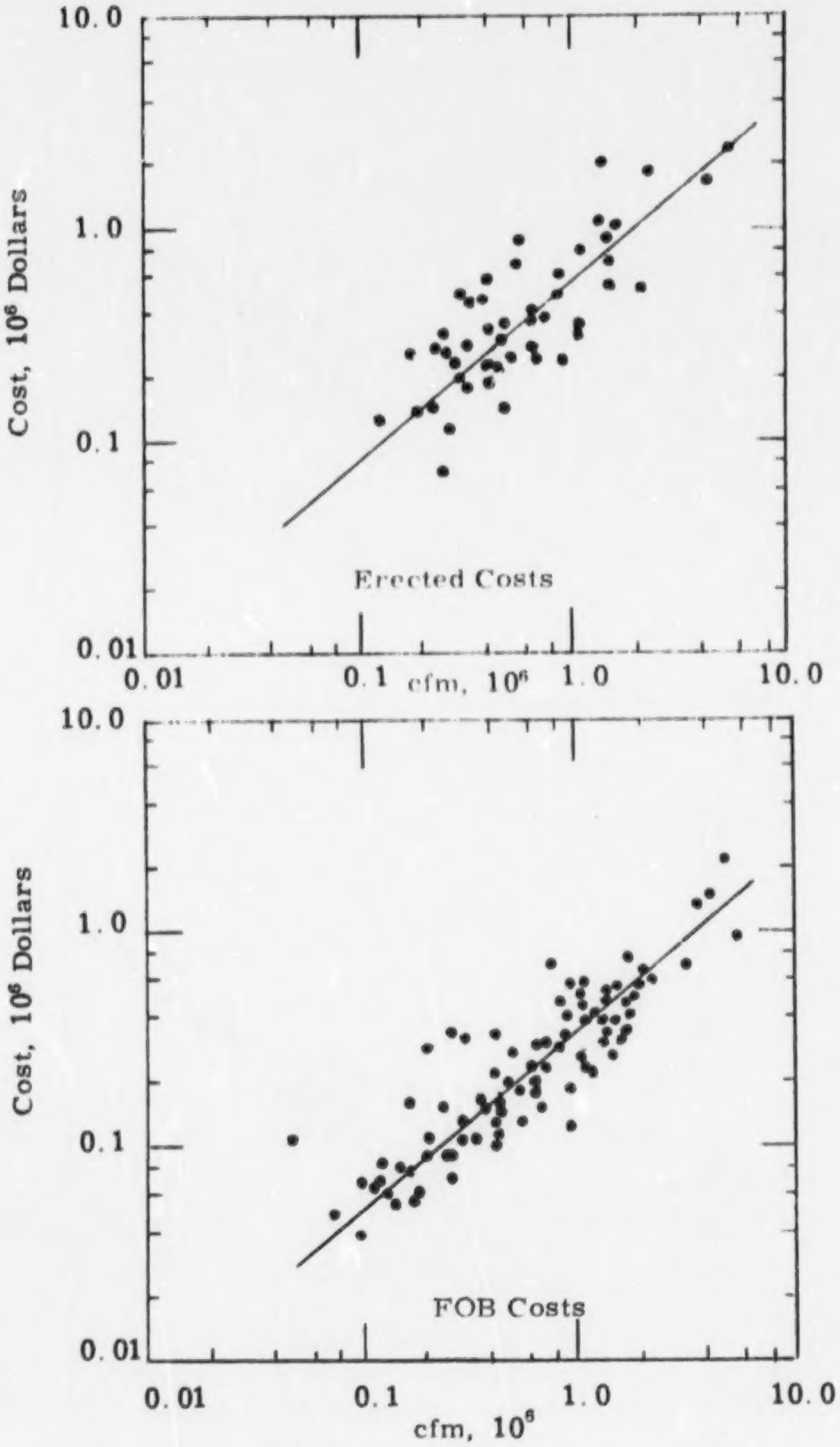

Figure 15.26. Electric Utility Electrostatic Precipitator Costs Vs. Volume Flow for Collection Efficiencies between 95 and 99%. These Costs Represent Installations between 1959 and 1969. The Costs Were Corrected to 1963 Dollars.

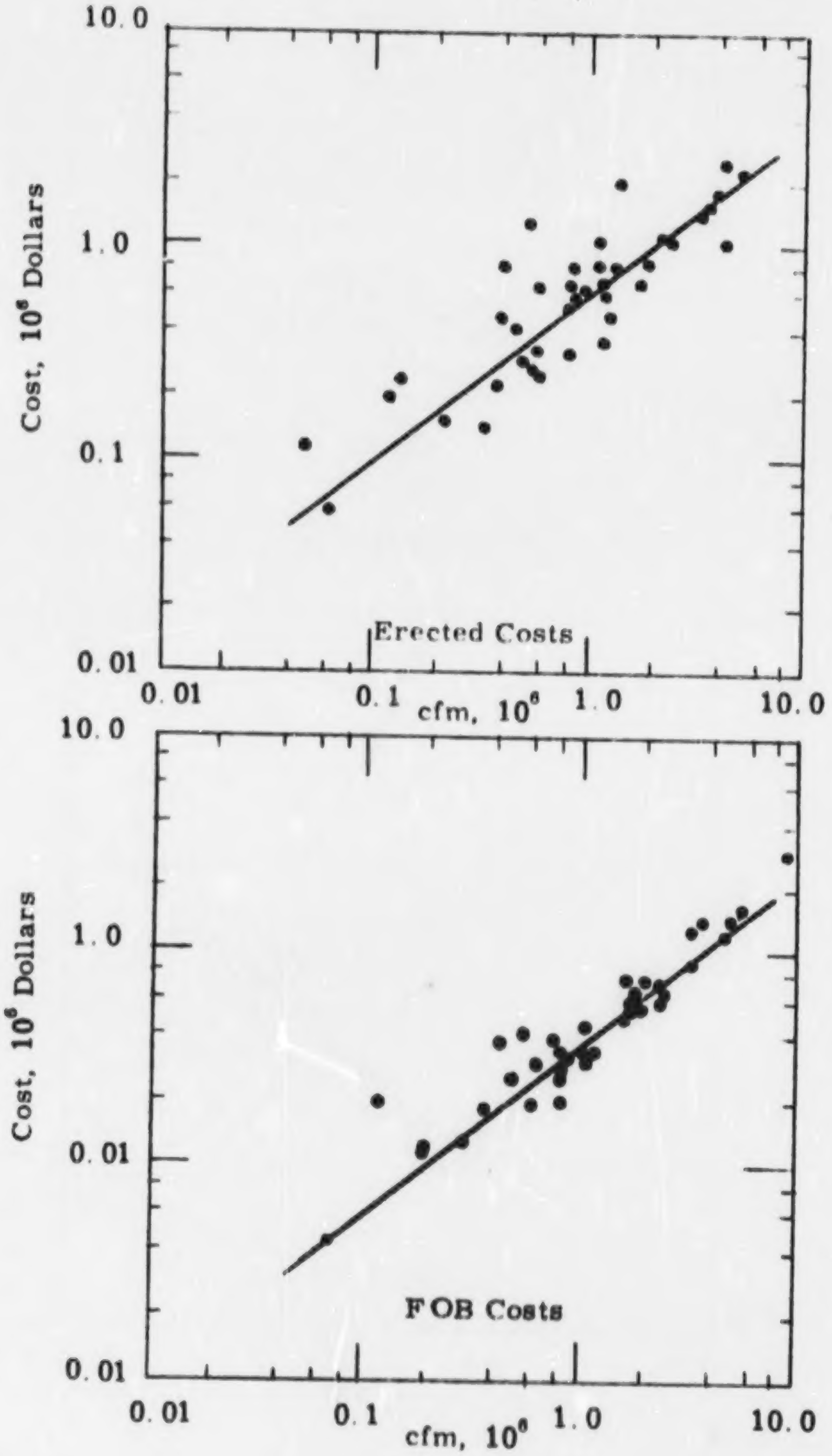

Figure 15.27. Electric Utility Electrostatic Precipitator Costs Vs. Volume Flow for Collection Efficiencies Greater Than 99%. These Costs Represent Installations between 1959 and 1969. The Costs Were Corrected to 1963 Dollars.

Table 15.7

Average Fly Ash Precipitator Costs from 1923 through 1969

	Number of Installations	FOB	Number of Installations	Erected
1923 - 1929	-	No data	9	22.8
1930 - 1934	-	No data	13	29.1
1935 - 1939	-	No data	39	26.3
1940 - 1944	-	No data	70	25.3
1945 - 1949	-	No data	87	39.5
1950 - 1954	65	28.2	64	50.1
1955 - 1959	-	No data	34	51.6
1960 - 1964	39	36.8	22	62.1
1965 - 1969	99	47.7	87	80.5

Note: Figures are an overall average for efficiency range from 90 to 99+% and gas volume range from $<10^5$ to $>10^6$ acfm.

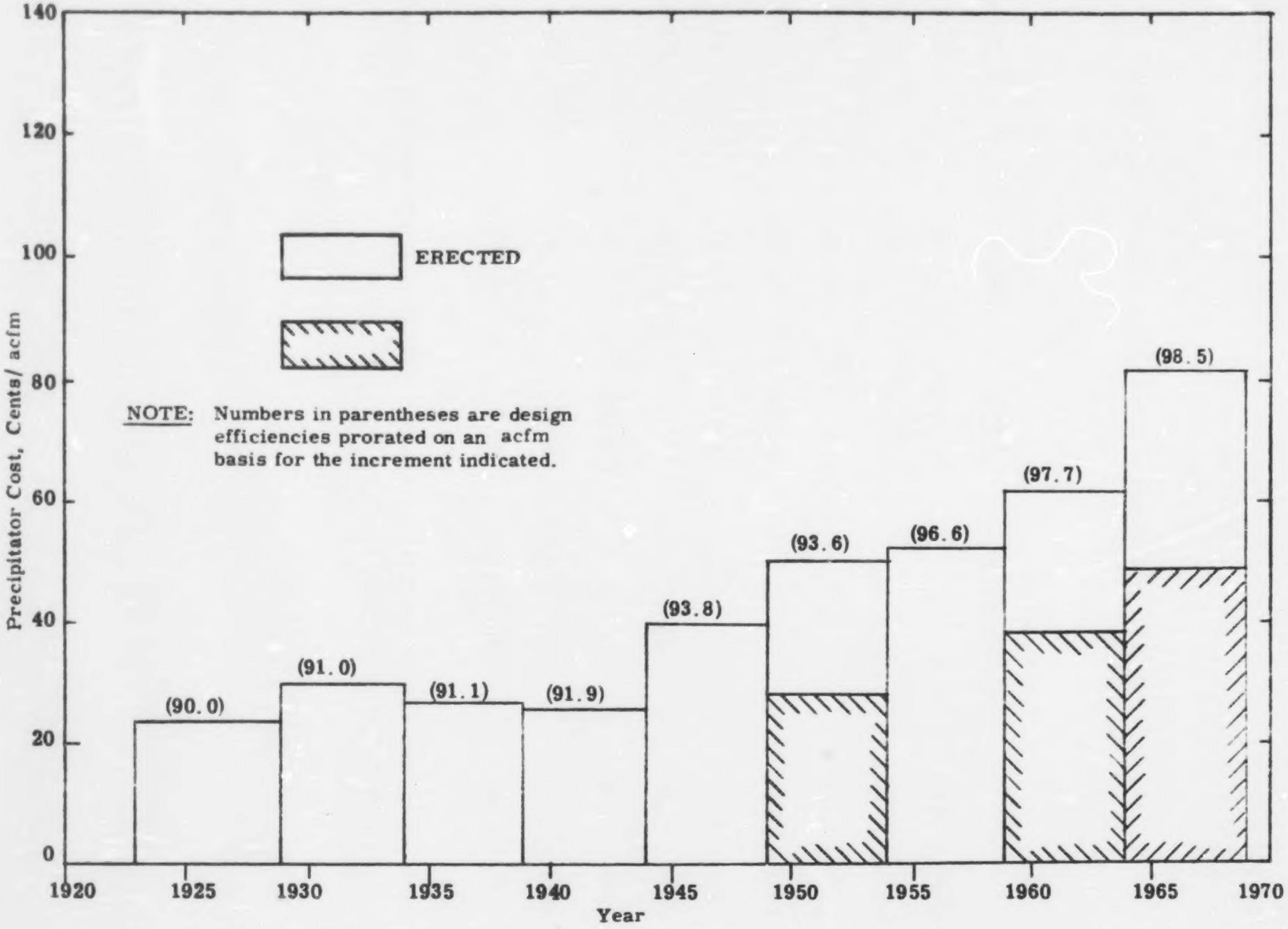

Figure 15.28. Precipitator Costs (Average) Over the Period 1923 through 1969.

Maintenance and operating costs. Maintenance and operating costs are difficult to establish since these costs are often not kept separately from other plant costs. Maintenance costs can vary from plant to plant depending upon the extent to which the plant is maintained. This in turn is dependent to a large degree upon company policy and the extent to which pressure is applied to maintain emissions within prescribed limits.

Tables 15.8 to 15.11 list representative operating costs for precipitators handling 200, 000, 500, 000, 1, 000, 000, and 2, 000, 000 cfm of gas. Table 15.12 is a summary of the data giving average operating costs for each size precipitator. The operating costs are made up of electric power, maintenance, operating, and premium charges.

Electric power costs were estimated from the installed electric power on those installations where actual costs were not available. Maintenance costs were estimated from a graduated scale depending upon the size of the units on those size installations for which actual maintenance costs were not available. Data by Van Huffel[15] and Waker[16] were used as a basis for these estimated costs.

Operating labor is not kept as a separate item of costs by most utilities. No operating personnel is required for an electrostatic precipitator on a full time basis. However, some tasks, such as routine meter reading and log keeping, are required and these are normally handled by regular plant personnel on a part time basis. In computing costs, operating labor was included as an estimate of the portion of time spent in these tasks. Premium charges were computed on the basis of 12.8% of the total installed precipitator costs. This item covers interest and depreciation.

Footnotes for Tables 15.8 to 15.11 give the basis for estimating the costs for the various items. Entries in the table with no footnotes are those obtained from company records.

15.7 TRENDS

A review of the electrostatic precipitator installations since 1945 shows that the installed volume has increased steadily over this period from a 5 million acfm yearly average in 1945 to about 16 million acfm

Table 15.8

Summary of Operating Costs for Fly Ash Precipitators

Gas Flow Rate	200,000 acfm	500,000 acfm	1 million acfm	2 million acfm
Electric Power	$ 2,075	$ 7,112	$ 11,500	$ 17,962
Maintenance	4,250	10,900	12,500	20,000
Operating Labor	2,400	3,200	4,500	6,000
Direct Operating Cost	$ 8,725	$21,212	$ 28,500	$ 43,962
Premium Charges	33,100	65,375	116,250	239,750
Total Charges	$41,825	$86,587	$144,700	$283,712

Table 15.9

Operating Costs for Precipitators for Electric Utilities

Gas Flow Rate = 200,000 acfm

	No. 1	No. 2	No. 3	No. 4	Average
Electric Power	$ 685	$ 1,810	$ 3,800[1]	$ 2,000[1]	$ 2,075
Maintenance	2,625	2,380	6,000[2]	6,000[2]	4,250
Operating Labor[1]	2,400	2,400	2,400	2,400	2,400
Direct Operating Cost	$ 5,710	$ 6,590	$12,200	$10,400	$ 8,725
Premium Charges[4]	34,500	34,500	32,000	31,400	33,100
Total Charges	$40,210	$41,090	$44,200	$41,800	$41,825

[1] Power cost estimated from installed capacity at $0.005/kWh.
[2] Maintenance cost estimated from Van Huffel and Walker.
[3] Operating labor estimated at 360 man-hrs per year at $6 per hr.
[4] Premium charge of 12.8% of capital cost for depreciation and capital charges.

Table 15.10

Operating Costs for Precipitators for Electric Utilities

Gas Flow Rate = 500, 000 acfm

	No. 1	No. 2	No. 3	No. 4	Average
Electric Power	$ 7, 200[1]	$ 8, 000[1]	$ 7, 000[1]	$ 6, 250	$ 7, 112
Maintenance	10, 000[2]	8, 550[2]	10, 000	15, 000	10, 900
Operating Labor[3]	3, 200	3, 200	3, 200	3, 200	3, 200
Direct Operating Cost	$20, 400	$19, 750	$20, 200	$ 24, 450	$21, 212
Premium Charges[4]	64, 000	43, 500	64, 000	90, 000	65, 375
Total Charges	$84, 400	$63, 250	$84, 200	$114, 450	$86, 587

[1] Power cost estimated from installed capacity at $0. 005/kWh.

[2] Maintenance cost estimated from Van Huffel and Walker.

[3] Operating labor estimated at 480 man-hrs per year at $6 per hr.

[4] Premium charge of 12.8% of capital cost for depreciation and capital charges.

Table 15.11

Operating Costs for Precipitators for Electric Utilities

Gas Flow Rate = 1 million acfm

	No. 1	No. 2	No. 3	No. 4	Average
Electric Power	$ 19,800[1]	$ 15,200[1]	$ 5,500	$ 5,400	$ 11,500
Maintenance	13,350[2]	13,950[2]	11,100	11,900	12,500
Operating Labor[3]	4,500	4,500	4,500	4,500	4,500
Direct Operating Cost	$ 37,650	$ 33,650	$ 21,100	$ 21,800	$ 28,500
Premium Charges[4]	100,000	110,000	128,000	128,000	116,250
Total Charges	$137,650	$143,650	$149,000	$149,800	$144,700

[1] Power cost estimated from installed capacity at $0.005/kWh.
[2] Maintenance cost estimated from Van Huffel and Walker.
[3] Operating labor estimated at 675 man-hrs per year at $6 per hr.
[4] Premium charge of 12.8% of capital cost for depreciation and capital charges.

Table 15.12

Operating Costs for Precipitators for Electric Utilities

Gas Flow Rate = 2 million acfm

	No. 1	No. 2	No. 3	No. 4	Average
Electric Power[1]	$ 11, 250	$ 10, 600	$ 30, 000	$ 20, 000	$ 17, 962
Maintenance[2]	20, 000	20, 000	20, 000	20, 000	20, 000
Operating Labor[3]	6, 000	6, 000	6, 000	6, 000	6, 000
Direct Operating Cost	$ 37, 250	$ 36, 600	$ 56, 000	$ 46, 000	$ 43, 962
Premium Charges[4]	192, 000	196, 000	251, 000	320, 000	239, 750
Total Charges	$229, 250	$232, 600	$307, 000	$366, 000	$283, 712

[1] Power cost estimated from installed capacity at $0. 005/kWh.
[2] Maintenance cost estimated at $0. 01/acfm per year.
[3] Operating labor estimated at 900 man-hrs per year at $6 per hr.
[4] Premium charge of 12. 8% of capital cost for depreciation and capital charges.

in the period 1955 to 1965. From 1965-1969, there has been a sharp increase to about 50 million acfm average per year. The increase in precipitator capacity can be attributed to the increased emphasis in air pollution control caused by the enactment of the clean air act and the subsequent establishment of particulate emission criteria. These have resulted in the installation of higher performance precipitators as back fits or up grading equipment to meet more stringent codes. Another factor in the growth has been the increased electrical energy demands, which have grown at a rate of 8-10% per year. Although the growth of nuclear power was predicted to provide a greater portion of the generating capacity, the construction of nuclear plants has been slowed by various factors and coal fueled steam plants have been used to meet the demand.

In addition to the potential application of nuclear plants, there are several alternative operating cycles for generating electricity with coal in large central-station power plants. The new operating cycles are in various stages of research and development at this point in time. Those which appear to be reasonably well along in the solution of technical problems and should be ready for use in the next decade are: the high-temperature steam cycle; the high-temperature gas turbine-steam turbine combined cycle; the supercharged boiler combined cycle; and the MHD-steam combined cycle. Several of these schemes require high efficiency ash removal systems operating at 1500-1700° F, and pressures in the order of 6 atmospheres (90 psia).

Techniques for removal of sulfur oxides from power plant effluents can drastically alter the performance of precipitators as well as the use of precipitators for fly ash collection. At this juncture, no decision has been reached as to the method or methods of sulfur oxide emission control that will be used.

Removal of the iron pyrites from coal is one technique for reducing sulfur oxide emissions. This process would in effect lower the sulfur content of the coal and in most instances would result in higher fly ash resistivities than would the untreated coal. Higher resistivity would tend to lower the precipitation rate parameter and require a greater collecting surface area to achieve a given efficiency.

Injection of limestone into the boiler to react with the sulfur oxide is also being studied as a means of controlling sulfur oxide emissions. Tests indicate that the limestone injection doubles the dust burden to the precipitator and substantially increases the resistivity. The higher resistivity dust would again increase the size and cost of a precipitator to achieve a given efficiency.

Two processes being considered, catalytic oxidation and molten salt scrubbers, require cleaning of the gas before it contacts the catalyst or prior to entering the scrubber. High temperature precipitators 1700° F) would be suited to this type service.

There is considerable interest in the use of high energy scrubbers for removal of SO_2 from the flue gas, and there are indications that particulate removal can be achieved along with the removal of the sulfur oxides. If such proves to be feasible, wet scrubbers might replace electrostatic precipitators for some applications. However, the large energy requirements necessary to handle the large gas flows, the requirement for reheating to restore plume bouyancy, the increased corrosion potential, and the water pollution problems created would tend to put wet scrubbers in less favorable position.

Other possibilities of control of sulfur oxides by fuel processing by methods such as solvent extraction, low temperature carbonization, or liquification, would alter the requirements for particulate control and would influence both the application and performance of precipitators.

CHAPTER 15
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CHAPTER 16

ELECTROSTATIC PRECIPITATORS IN THE PULP AND PAPER INDUSTRY

Electrostatic precipitators are used in the pulp and paper industry primarily to clean flue gases emitted by recovery boilers. These boilers are installed in all modern mills that use significant amounts of chemicals in the pulp-making process as a part of the recovery cycle. The development and application of this cycle, which permits recovery and reuse of chemical additives, has been instrumental in making chemical pulp processing economically feasible. The electrostatic precipitator is an integral part of the chemical recovery cycle of the pulp and paper mill, and its application has grown largely in proportion to the growth of the chemical pulping industry. It appears that this growth trend will continue and perhaps accelerate in the future.

Virtually all paper and paper products manufactured in the United States are made from wood pulp. There are two basic techniques available for reducing the logs to the pulp feed to the paper machine. These are: (1) mechanical means that use large grindstones to shred the logs and groundwood refiners that produce mechanical pulp from chips, and (2) chemical methods in which the organic materials are dissolved to release and separate the useful cellulose fibers which comprise about 50% of the raw wood. Chemical methods usually require reduction of the logs to chips by mechanical means prior to chemical processing.

Many combinations and variations of these two pulping techniques have been developed, including one common method consisting of a mild chemical pre-treatment stage to soften the chips followed by mechanical action to separate the individual fibers. This process is known as the semichemical or chemimechanical method.

A summary of the basic pulping processes in common usage is given in Table 16.1. The separation into three columns reflects the yield or approximate precentage of raw wood that is made into pulp. It will be seen that mechanical methods, which produce a less refined pulp, have the greatest yield, and chemical methods the least, with the semichemical in between.

Table 16.1

Summary of Basic Pulping Processes

	Mechanical	Semichemical	Chemical
A. Yield	90%+	70-85%	40-60%
B. Basic Process	Fibers are separated by mechanical means, using: 1. Artificial stone grinders 2. Revolving disc grinders 3. Expansion guns 4. Others (rare)	Two-stage process: 1. Chemical pretreatment (see below) followed by: 2. Mechanical fiber separation (see left-hand column) (First stage is mild application of any typical chemical process --see list in right-hand column--neutral sulfite most common, cold soda also used.)	Fibers are separated by chemical action at elevated pressure and temperature; lignin bonds between fibers are dissolved. For yields in 50-60% range, the chemical treatment is decreased and the pulp is refined by mechanical means. Principal chemical processes are: 1. Sulfite (acid, neutral, or alkaline) 2. Sulfate (alkaline) 3. Soda (alkaline)
C. Typical Products	1. Newsprint 2. Publication grade (magazine, book, etc.) 3. Wall paper 4. Insulation board 5. Building board	1. Corrugating medium for paperboard boxes 2. Newsprint (partial replacement for mechanical pulp)	1. Linerboard 2. Newsprint (30% Kraft) 3. Publication grade 4. Book papers 5. Tissue 6. Towelling 7. Bag 8. Specialty types

The principal chemical processes are sulfite, sulfate or kraft, and soda, taking their names from the active constituents of the cooking liquor. The liquor is applied to the wood chips in a steam-heated pressure vessel under carefully controlled conditions of temperature, pressure, and cooking time. Cooking dissolves the organic binder materials in the chips, releasing the cellulose fibers. Any of these methods can be used to soften the wood in the semichemical process by reducing the cooking time, temperature, or amount of chemical charge and using mechanical means to separate the fibers. The sulfate and soda processes are always alkaline; the sulfite process can vary from highly acid to highly alkaline (the latter case is rather rare, however).

Any of these pulping methods, mechanical, semichemical, or chemical, can have a bleaching stage to whiten the product. Bleaching is applied in varying degrees to make pure white papers such as fine writing and printing papers to the somewhat off-white colors of newsprint and pulp-type magazine papers. The sulfate process, when unbleached, produces the typical brown-colored wrapping, bag, and cardboard carton stock in everyday use.

A recent survey of chemical and semichemical mills in the United States showed the following breakdown by process usage.

Table 16.2

A Survey of Chemical and Semichemical Mills in the United States

	No. of Mills	Percentage
Sulfate or Kraft (SO_4)	124	55.5
Semichemical	48	21.5
Sulfite (SO_3)	44	19.3
Soda	3	1.4
Other	5	2.3
	224	100.0

The mills listed as "other" represent a unique variation of the sulfite process in which magnesium is used as the base chemical instead of the more common sodium, calcium, or ammonium radical.

The kraft process is a modification of the soda process in which the cooking liquor is a mixture of sodium hydroxide, sodium carbonate, sodium sulfide, and sodium sulfate; the make-up sulfur is supplied by additions of sodium sulfate (salt cake), which is reduced to sulfide in the recovery cycle. This process has an economic advantage over the soda process in that it yields a stronger pulp and is not limited in the wood species that can be pulped. Production growth of the kraft process since 1940 has continued at a rapid pace, reaching nearly thirty million tons per year in 1965. Figure 16.1 shows pulp production in the United States by principal processes from 1900 to 1965.[1,2]

The economics of the kraft process require recovery and reuse of the spent cooking chemicals and recovery of the heat content of the concentrated spent liquor. This chemical and energy recovery is accomplished by combustion, which generates large volumes of flue gases. During combustion, a significant fraction of the recoverable chemicals is entrained as particulate matter with the flue gas; hence the widespread use of particulate collection devices. In view of the fact that application of electrostatic precipitators in the pulp and paper industry has been almost exclusively to sulfate pulp mills, this section will be limited to the sulfate process, with only a brief survey of the other processes.

16.1 PROCESS DESCRIPTIONS

A summary of the chemical cooking processes is given in Table 16.3. It gives the process name, active ingredients in the cooking liquor, and approximate pH range of the liquor as originally introduced into the vessel that cooks the chips under conditions of elevated temperature and pressure. Note that the cooking process causes various chemical reactions as the liquors dissolve the organic binders and that, in general, these reactions tend to change the pH as the cooking proceeds.

Soda process. Sodium hydroxide solution under pressure is the cooking liquor in this process, which took its name from the sodium carbonate used as make-up chemical. From the 1850's until around 1900, this process was the principal source of wood pulp in the United States and production continued to grow until about 1920, but has remained essentially static since then. This process never exceeded a production of about one-half million tons per year.

Sulfite process (SO_3). The sulfite process has four principal variations in which the cooking liquor varies from highly acid to highly alkaline. The acid sulfite process has both sulfurous acid and a bisulfite compound in the cooking liquor. This gives it the characteristic low pH. The bisulfite method uses only the bisulfite compound, commonly sodium, calcium, or ammonium. Where calcium is used as the base, an excess of SO_2 is

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[1] Refer to the bibliography for this chapter.

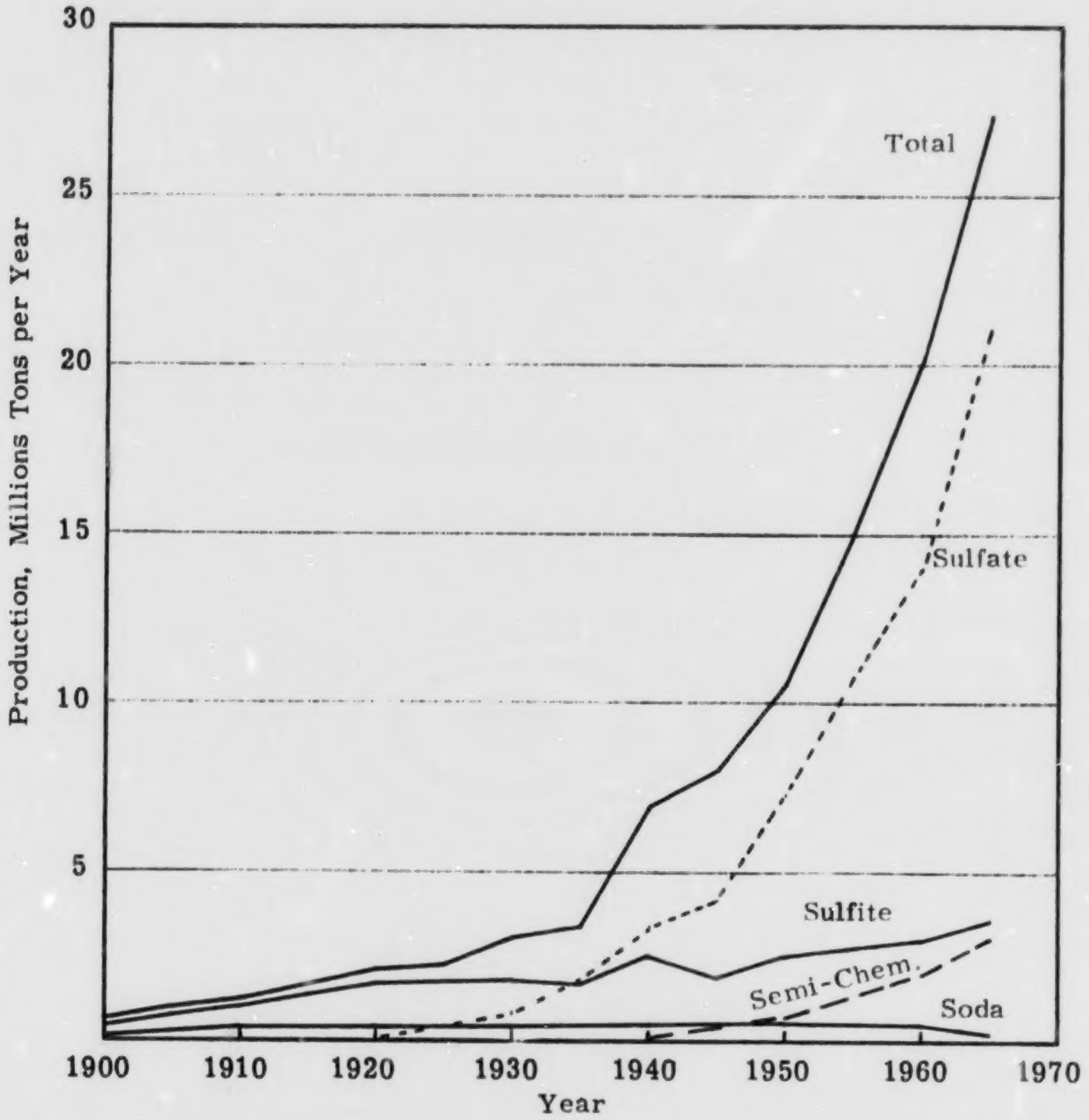

Figure 16.1. United States Production of Chemical and Semi-Chemical Pulps (1900-1965).

Table 16.3

Summary of Chemical Cooking Processes

Process Name	Active Ingredients in Cooking Liquor	pH of Cooking Liquor
Acid Sulfite	H_2SO_3 + (X) HSO_3	1 - 2
Bisulfite	(X) HSO_3	2 - 6
Neutral Sulfite	(X) SO_3 + (X) CO_3 (may have (X) OH)	6 - 9
Alkaline Sulfite (rarely used)	(X) SO_3 + (X) OH (or Na_2S)	10+
Sulfate (kraft)	NaOH + Na_2S + Na_2CO_3	13 - 14
Soda (rarely used)	NaOH	14

X = Ca, Mg, Na, or NH_3

required in the cooking liquor to maintain solubility of the calcium bisulfite at elevated temperatures. Other bases, such as sodium and ammonium, required little or no excess SO_2, hence they can have higher pH values. Magnesium bisulfite is used in a few mills, and termed the magnefite or MgO process. Neutral sulfite includes a bicarbonate as a buffering compound to neutralize the acidity of the bisulfite; this process is usually neutral or slightly alkaline. The alkaline sulfite method with a sulfide or hydroxide in the cooking liquor is rarely used.

The first commercial pulp mills using this process were built in the United States during the 1880's and production from this process expanded much more rapidly than the soda process. For about thirty years, starting around 1900, the sulfite process dominated the chemical production of wood pulp in this country, reaching a level of about three to four million tons per year. Since about 1940, growth in production has been only slight.

Kraft pulping process (SO_4). In brief, this process consists of chemical and mechanical separation of the main components of wood; cellulose and lignin. The cellulose pulp is recovered as the product; the chemicals used in dissolving the lignin are recovered for reuse; and the other components are recovered as by-products or burned as fuel in the process. Losses from the system include some of the reaction chemicals and some of the non-cellulose components; these are discharged as waste in water or to the atmosphere at various steps in the process.

Figure 16. 2 is a simplified flow sheet for the production of pulp by the kraft method.

Referring to Figure 16. 2, it can be seen that the kraft process is composed of several operations which include chemical digestion, filtration, washing, evaporation, black liquor recovery, and causticizing.

In the chemical digestion process, wood chips are cooked in the digesters with a solution of sodium hydroxide and sodium sulfide. The chips and the cooking solution, called white liquor, are heated to about 340°F and maintained at 100 psi for approximately 3 hours. During this time, hydrolysis of lignin and carbohydrates occurs, fats and resins are saponified and made soluble, and the cellulose fibers (pulp) are left free and in suspension. In modern mills, digestion takes place in a continuous process.

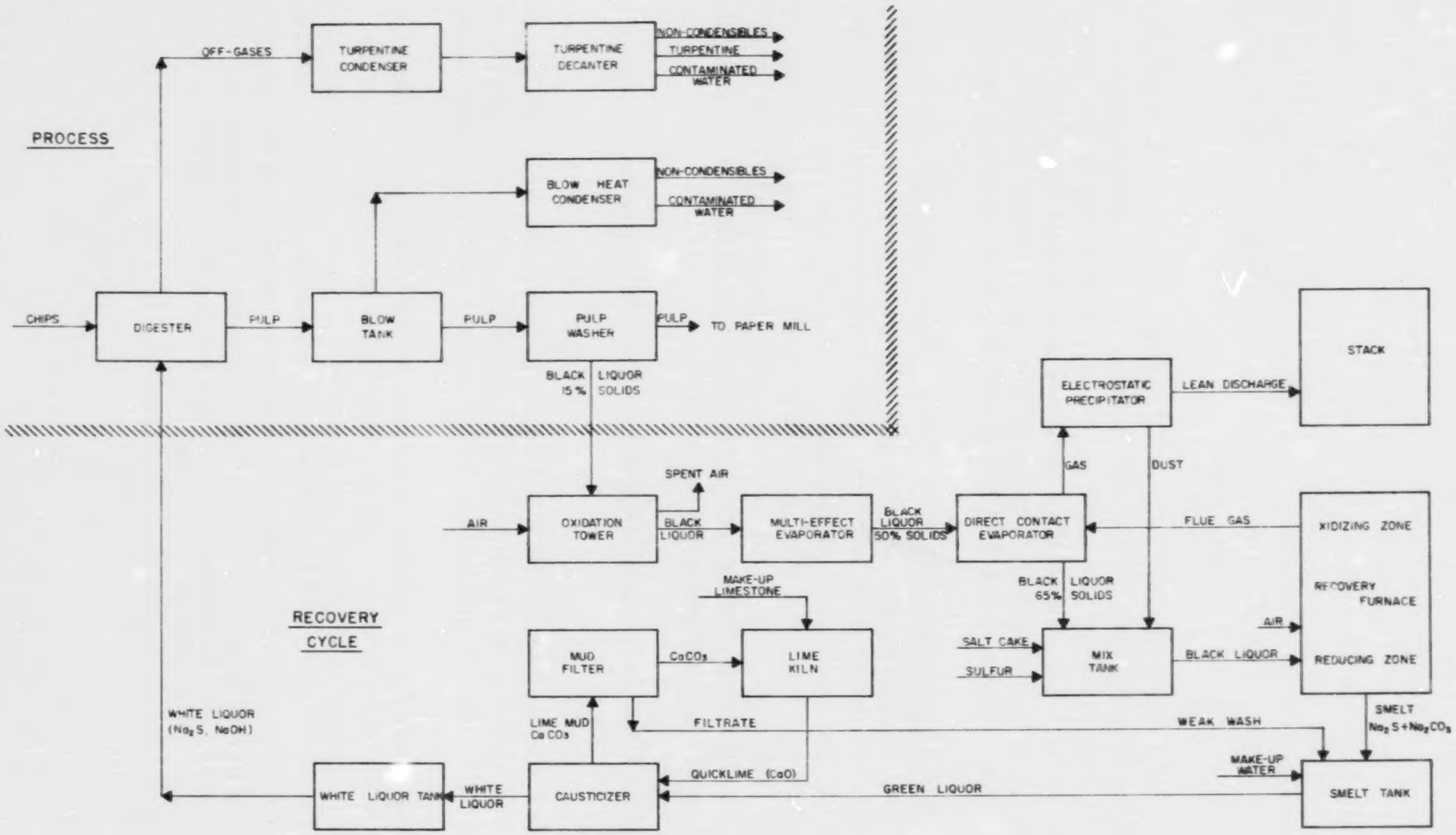
PROCESS
CHIPS
DIGESTER
OFF-GASES
TURPENTINE CONDENSER
TURPENTINE DECANTER
NON-CONDENSIBLES
TURPENTINE
CONTAMINATED WATER
PULP
BLOW TANK
BLOW HEAT CONDENSER
NON-CONDENSIBLES
CONTAMINATED WATER
PULP
PULP WASHER
PULP
TO PAPER MILL
BLACK LIQUOR 15% SOLIDS
RECOVERY CYCLE
AIR
OXIDATION TOWER
SPENT AIR
BLACK LIQUOR
MULTI-EFFECT EVAPORATOR
BLACK LIQUOR 50% SOLIDS
DIRECT CONTACT EVAPORATOR
GAS
ELECTROSTATIC PRECIPITATOR
LEAN DISCHARGE
STACK
DUST
FLUE GAS
XIDIZING ZONE
RECOVERY FURNACE
REDUCING ZONE
AIR
BLACK LIQUOR 65% SOLIDS
SALT CAKE
SULFUR
MIX TANK
BLACK LIQUOR
SMELT $Na_2S+Na_2CO_3$
SMELT TANK
MAKE-UP WATER
WEAK WASH
GREEN LIQUOR
MAKE-UP LIMESTONE
LIME KILN
$CaCO_3$
MUD FILTER
FILTRATE
QUICKLIME (CaO)
LIME MUD $CaCO_3$
CAUSTICIZER
WHITE LIQUOR
WHITE LIQUOR TANK
WHITE LIQUOR (Na_2S, NaOH)

Figure 16.2. Simplified Flow Sheet, Kraft Pulping Process.

Following digestion, the pulp is separated from the cooking solution and washed. Washing is normally accomplished by countercurrent water flow on vacuum filters. A minimum amount of water is utilized since all of the water must be evaporated to recover the chemicals.

Filtrate from the washing operation, containing approximately 15% solids (black liquor), is processed for chemical recovery. While there are several variations in the evaporation-recovery operation, the following system is typical.

The first step in the recovery cycle is the concentration of the weak liquor from 15% solids to a viscous solution of about 50% solids in multiple effect, steam-heated evaporators. Beyond 50% concentration, it is evaporated by exposure to direct heat of the flue gases from the recovery furnace, and concentrated to a solids content of 62 to 65%.

Due to odor emission problems, there has been a design trend away from direct contact evaporation of black liquor with flue gases. This trend could have a significant influence on the electrostatic precipitation recovery systems for flue gases since the direct contact evaporator also acts as a wet scrubbing device.

The recovery operation plays an important role in the economics of the kraft process. The recovery furnace has two main functions; (1) to convert spent cooking chemicals to a reusable form, and (2) to recover the heating value of the concentrated spent liquor.

Concentrated black liquor is sprayed into the furnace where it is dehydrated by hot flue gases and the organic constituents burned. The liquor contains enough organic matter to burn and evaporate the remainder of the water, with some heat left over to produce steam in the furnace boiler. Depending on the particular design, the dehydration of the liquor takes place either during flight or after impact on the furnace tube walls. The ash collecting at the bottom part of the furnace is subjected to an oxygen deficient (reducing) atmosphere, and a smelt consisting chiefly of sodium carbonate and sodium sulfide is formed and continuously withdrawn to a smelt dissolving tank. The combustion of furnace gases is completed in the upper portion of the furnace by the introduction of secondary combustion air.

The combustion air creates a strong draft through the furnace and entrains appreciable quantities of particulate matter. The particulates consist mainly of sodium salts which are valuable enough to warrant recovery.

In most mills, particulate recovery is by electrostatic precipitation, and it is in this operation that most of the electrostatic precipitators utilized by the pulp and paper industry are found.

The main product of the furnace is a molten smelt consisting primarily of sodium carbonate and sodium sulfide, which is subsequently dissolved and diluted to the desired alkali concentration. This solution (green liquor) is then mixed with lime to convert the sodium carbonate to sodium hydroxide. At the completion of the step called causticizing, the solution (white liquor) containing sodium sulfide and sodium hydroxide is returned to the digester, and the precipitated calcium carbonate (limesludge) is calcined in a kiln and recycled. Particulate emission control on the lime kiln normally is accomplished by the use of wet scrubbing systems.

16.2 CHARACTER OF RECOVERY BOILER PARTICULATE EMISSION

Particle sampling and size distribution analysis. Because of the extreme fineness, hygroscopic nature, and impaction characteristics of the dust, it is difficult to collect a sample in the field and transport it to the laboratory for analysis. However, the following technique has been developed and used successfully.[3,4,5] A gas sample is drawn from the main flue system ahead of the precipitator through an insulated sampling probe at a high flow rate to prevent condensation of water vapor. The sample is introduced into a small, heated, point-plane electrostatic precipitator, in which the collecting electrode is a copper mesh electron microscope target. The precipitator is energized for several seconds, de-energized, and the target returned immediately to its capsule to protect it from moisture and mechanical effects. The precipitated dust on the target is magnified in an electron microscope and pictures made of the resulting image for particle sizing.

Photomicrographs of typical samples used in particle size and physical characteristic analyses are shown in Figures 16.3, 16.4, and 16.5. Figure 16.6 shows the count distribution of the sample in Figure 16.3, while Figure 16.7 is a logarithmic probability plot of a composite of several samples taken at the same location. These results show a log normal distribution, with a median diameter of 0.42μ (by number). As discussed later, flue gas conditions may exert pronounced effects on particle size and shape as illustrated by Figures 16.4 and 16.5, photomicrographs taken at different locations.

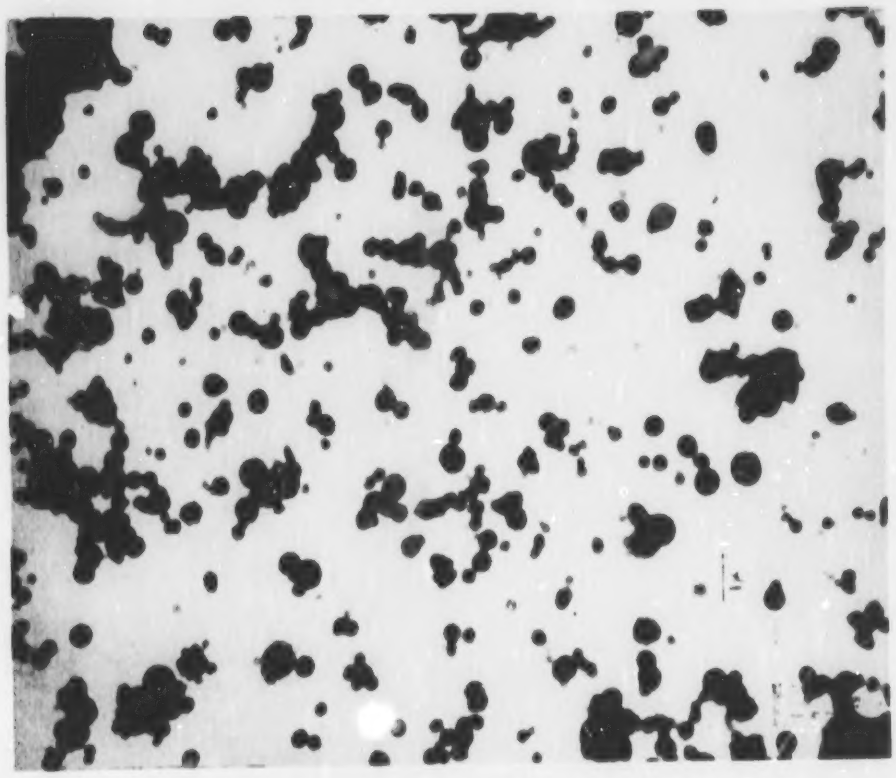

Figure 16.3. Temperature, Approximately 350°F.

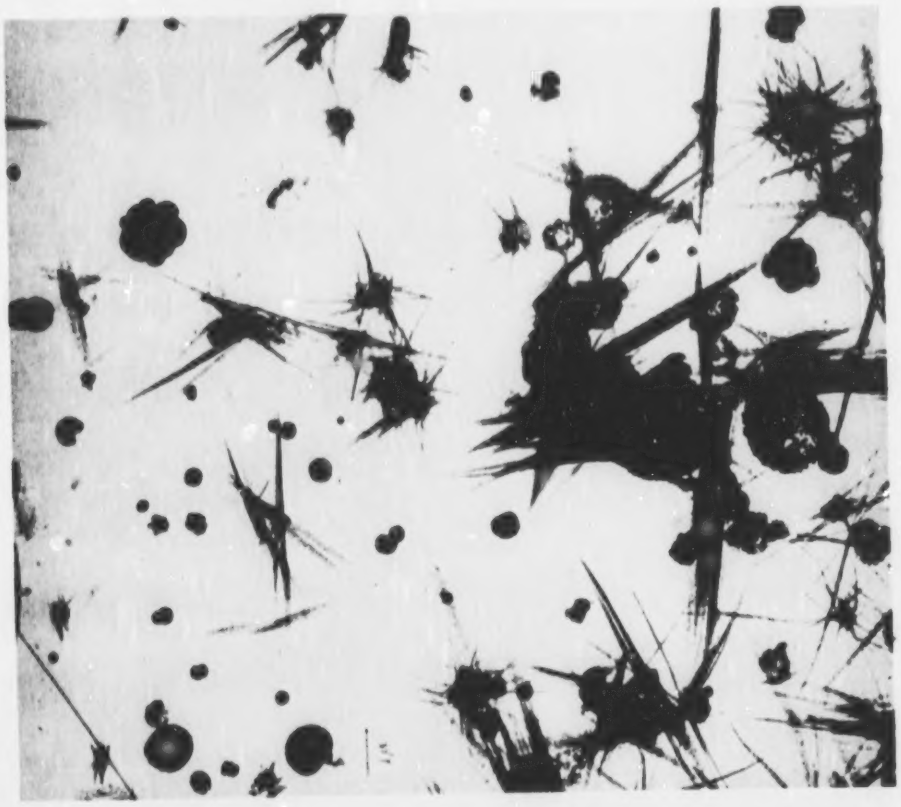

Figure 16.4. Temperature, Approximately 280°F.

Photomicrographs of Particles Collected from a Pulp Mill Recovery Unit

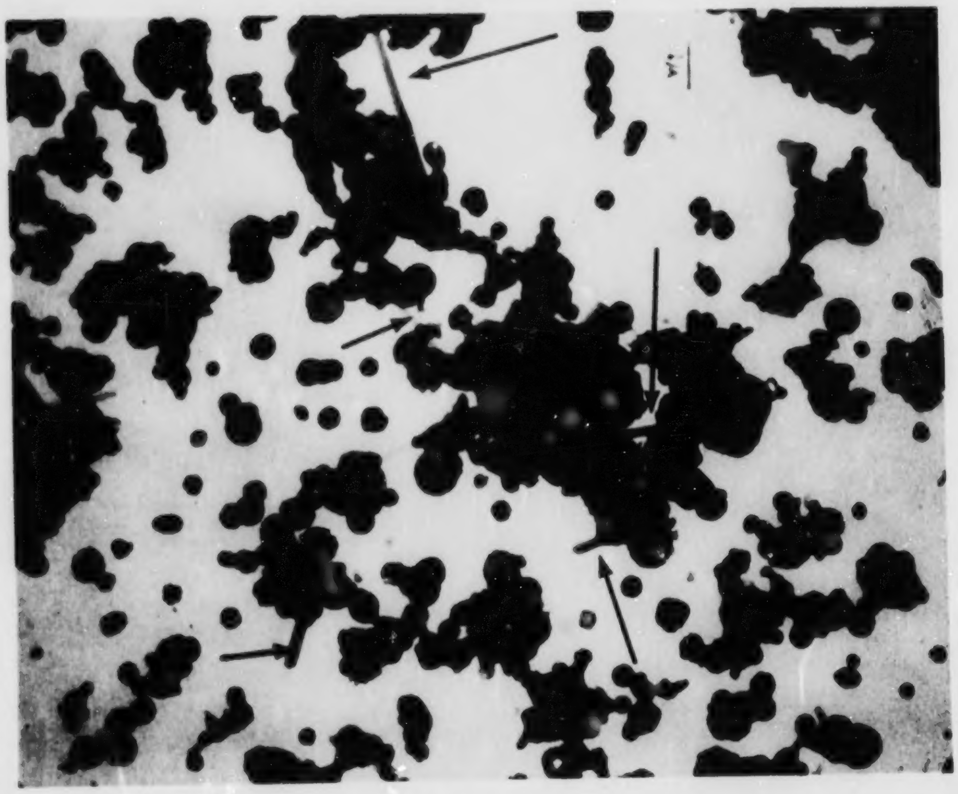

Figure 16.5. Photomicrograph of Particles Resulting from Mixing of Gases and Dust.

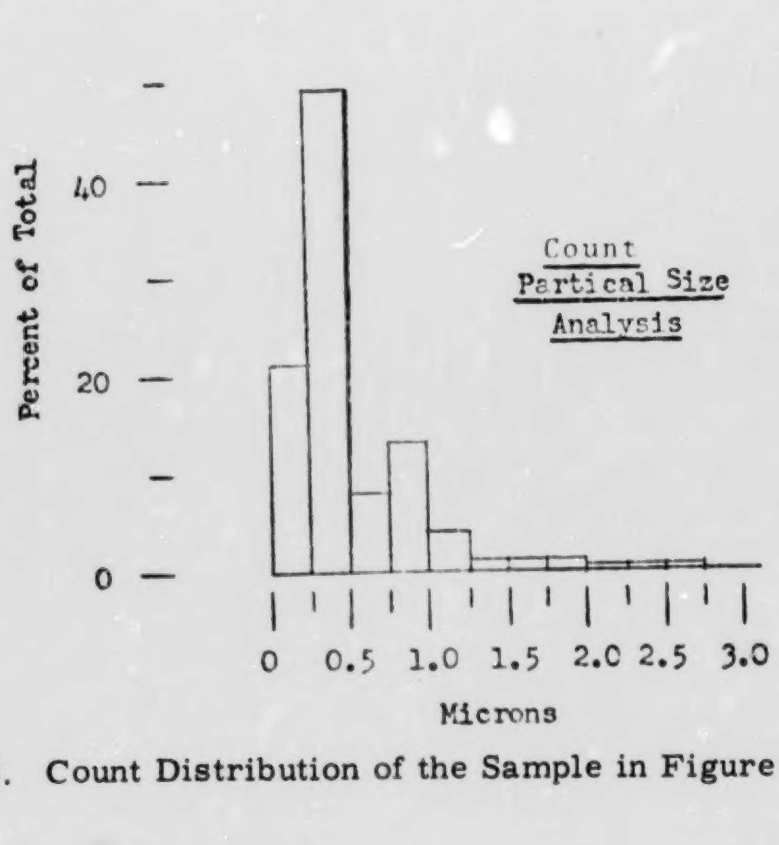

Figure 16.6. Count Distribution of the Sample in Figure 16.3.

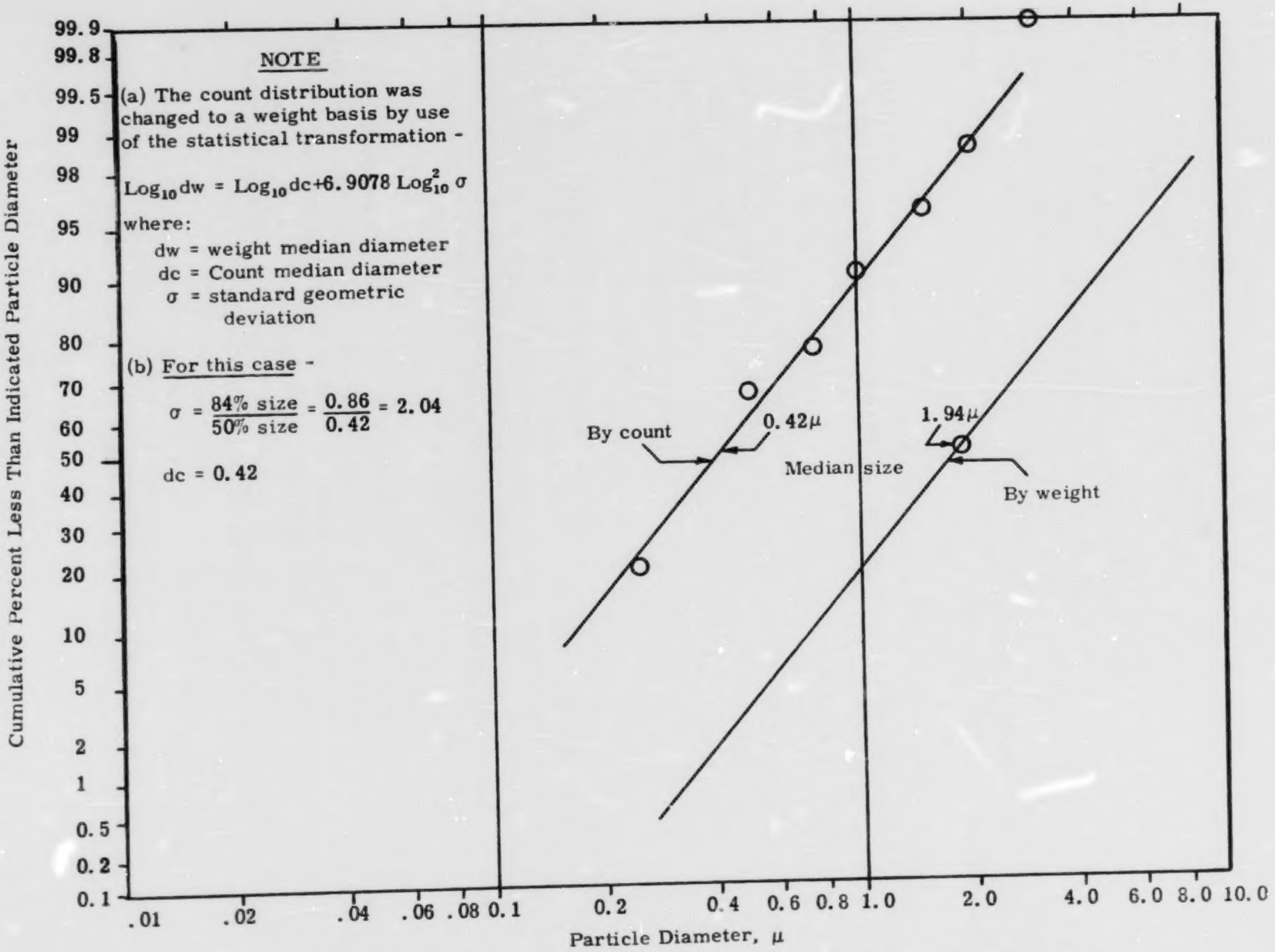

Figure 16.7. Particle Size Distribution of Recovery Furnace Fume.

Physical characteristics. These micrographs present an interesting interpretation of experience with paper mill dust characteristics. It has been noted that recovery furnace emissions appear to have a critical temperature characteristic. In some mills the dust is light, easily handled, transported in the flue systems, and easily rapped off the collecting electrodes in the precipitator, while other mills produce a sticky or heavy material which tends to build up in flue systems and is difficult to clean off the precipitator collecting electrodes. The pictures presented here illustrate this critical temperature concept.

Figure 16.3 represents a dust sample taken at a gas temperature of approximately 350-360°F, representing the solidification of the fume as suspended in the furnace gases. Note that the material is all in spherical form. Figure 16.4 is a sample taken at a lower temperature of approximately 280-290°F. It is conjectured that this temperature is below the hydration or adsorption point for particles of this fineness and chemical make-up, i.e., the particles act as "condensation nuclei" for the flue moisture which, at this particular plant, varied between 25 and 35% by volume. The particles hydrate and create the crystalline structure as shown.

Figure 16.5 is a sample taken from a flue after the gases and dust of Figures 16.3 and 16.4 had been thoroughly mixed. The resultant temperature increase to 310-320°F partially dehydrated the crystals of Figure 16.4. Several bars are seen and indicated by arrows in Figure 16.5. It is felt that these bars are the result of a rapid dehydration of the sharp points of the original crystals.

A logical conclusion is that the sticky, heavy material encountered at some plants is similar to that shown in Figure 16.4, while the light, dry material is like that shown in Figure 16.3. The critical temperature may vary from plant to plant depending upon the percent moisture in the gas, fineness of the dust, and its chemical characteristics. At this particular plant, the critical temperature appears to be about 310°F.

Chemical composition. The particulate emission from the recovery furnace in the sulfate process is comprised of alkali salts, principally sodium sulfate. A range of constituents found in a typical dust is as follows:

Constituent	Range (% by weight)
Na_2SO_4	75-90
Na_2CO_3	7-22
NaCl	1-1.5
Sulfides	0.5-1 (ASNa_2S)
Na_2SO_3	0.1-0.2

The dust as collected is extremely light in weight, averaging about 9 lb per cu ft when freshly precipitated, and about 30 lb per cu ft when packed.

Electrical resistivity. Figure 16.8 shows the electrical resistivity of salt cake (laboratory measurements). Results were obtained with only 1-2% moisture in the air. Figure 16.9 shows results of reagent grade sodium sulfate (anhydrous) for a range of moisture from 0 to 30% by volume.

There is little or no information available on the resistivity of salt cake measured in situ, and the data presented here only serve as a guide. In general, the moisture content of the exit gas from a recovery boiler is of the order of 20-40%, hence resistivity in the neighborhood of 10^9 ohm-cm could be expected.

No instances of difficulty attributable to high resistivity have been noted which, along with the difficulty in measurement, probably accounts for the lack of field resistivity measurements.

16.3 CHARACTER OF SULFATE PROCESS FLUE GASES

Typical characteristics of the inlet gas composition to the electrostatic precipitator are as follows:

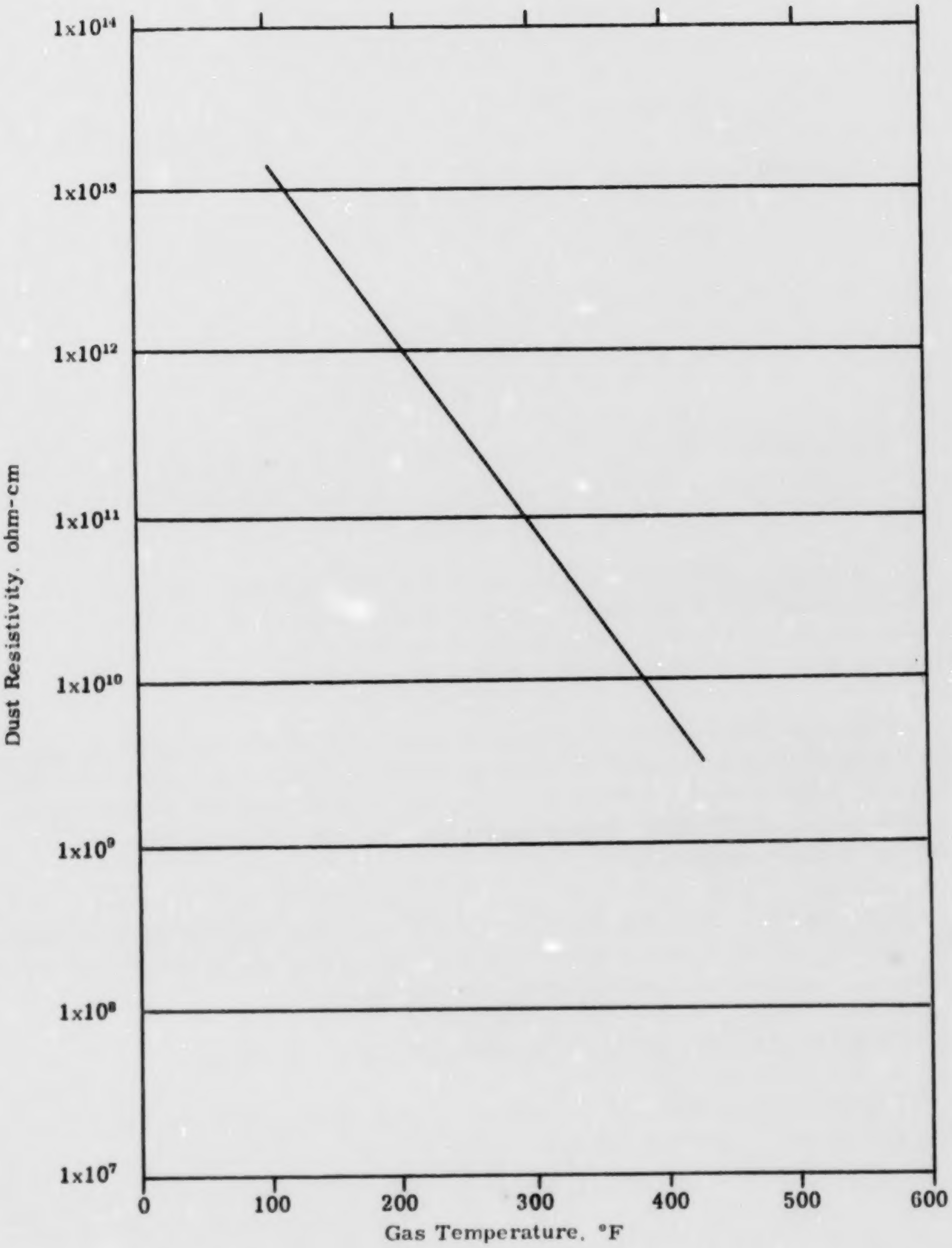

Figure **16.8.** Electrical Resistivity of Salt Cake (Laboratory Measurement with **1-2%** Moisture in Gas).

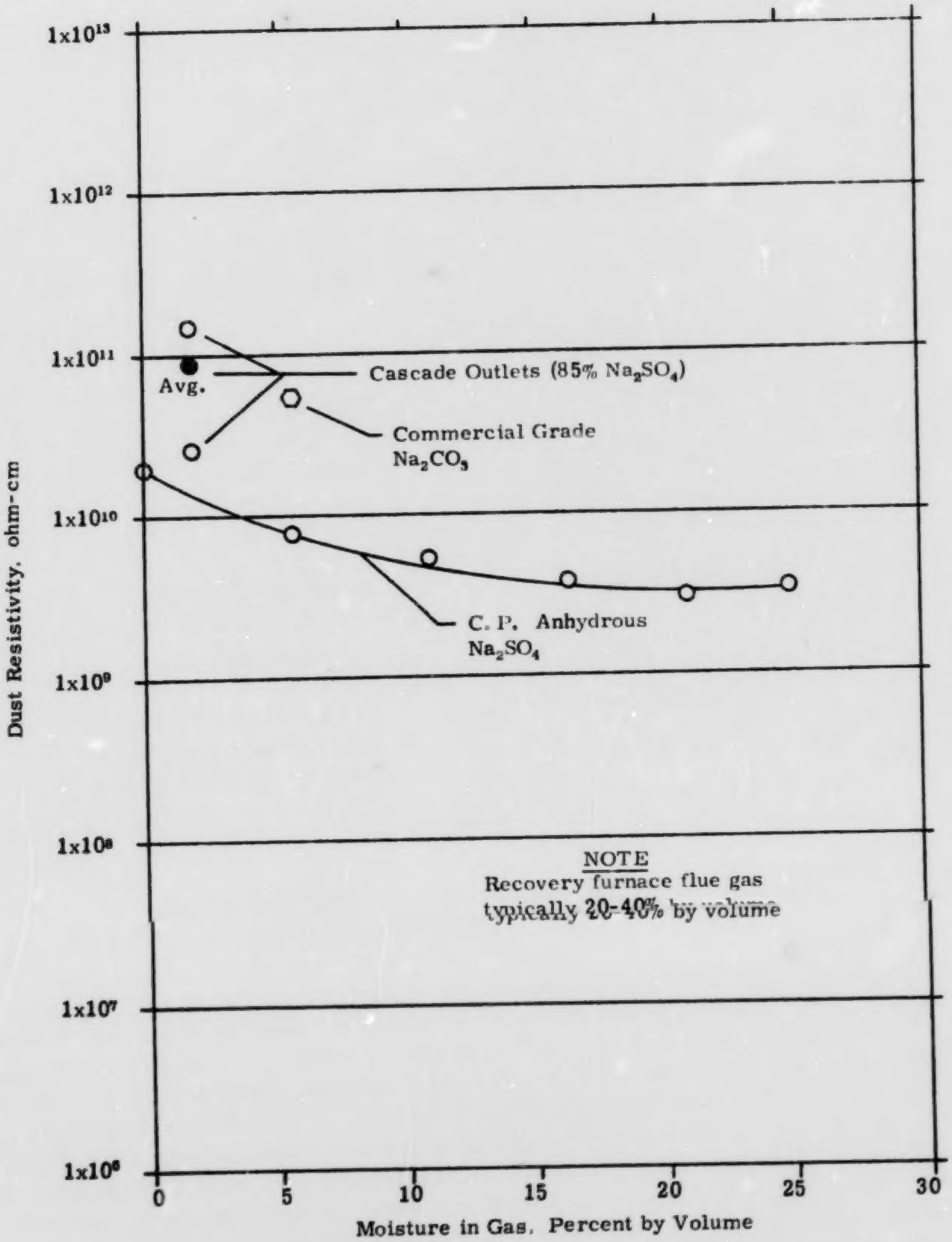

Figure 16.9. Electrical Resistivity of Sodium Sulfate as a Function of Moisture in Gas at 300°F (Laboratory Measurements).

Component	Volume % (dry)
N_2	81
CO_2	16.4
CO	0.1
O_2	2.5
Mercaptans	Trace
Sulfides	Trace
Temperature	260-320°F
Dewpoint	135-180°F
H_2O Content	20-40 (by volume)
Pressure	Atmospheric
Dust Concentration	1-9 grains/std cu ft of dry flue gas (typically, 3-6 grains/scfd at the precipitator inlet).

16.4 HISTORICAL DEVELOPMENTS OF ELECTROSTATIC PRECIPITATORS FOR RECOVERY FURNACES

The use of electrostatic precipitation for the collection of particle emissions from the recovery furnace commenced shortly after the development and adoption of the spray type recovery furnace and auxiliary evaporative equipment in the late 1930's. Following this commercial introduction in 1939, dry bottom electrostatic precipitators, illustrated in Figures 16.10 and 16.11, were utilized in all precipitator installations until 1948 when the "wet bottom" precipitator was introduced. In dry bottom precipitators, the collected dust is accumulated in dry pyramidal-type hoppers. The wet bottom precipitator, shown in Figures 16.12 and 16.13, was designed to avoid problems associated with collecting and redissolving

——— Indicates Inflowing Gas

- - - - - Indicates Outflowing Gas

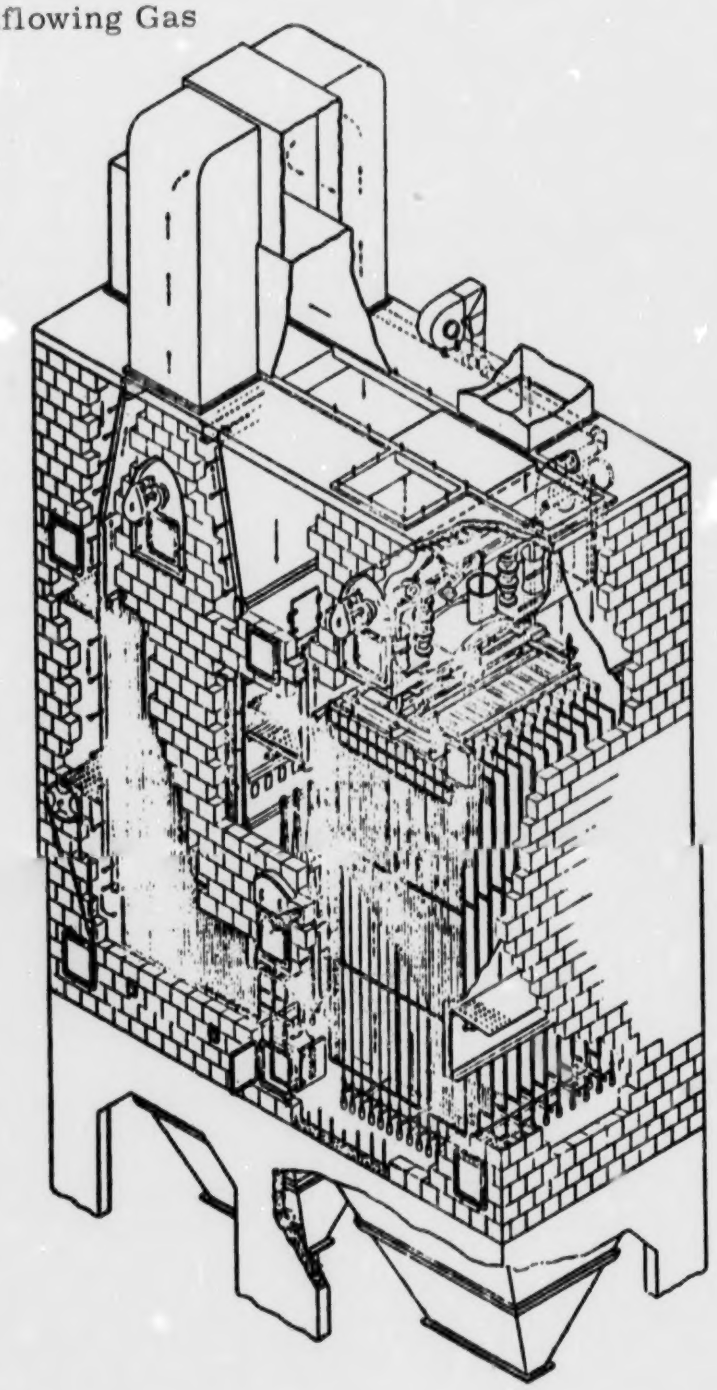

Figure 16.10. Vertical Flow Salt Cake Precipitator With Dry (Hopper) Dust Removal System.

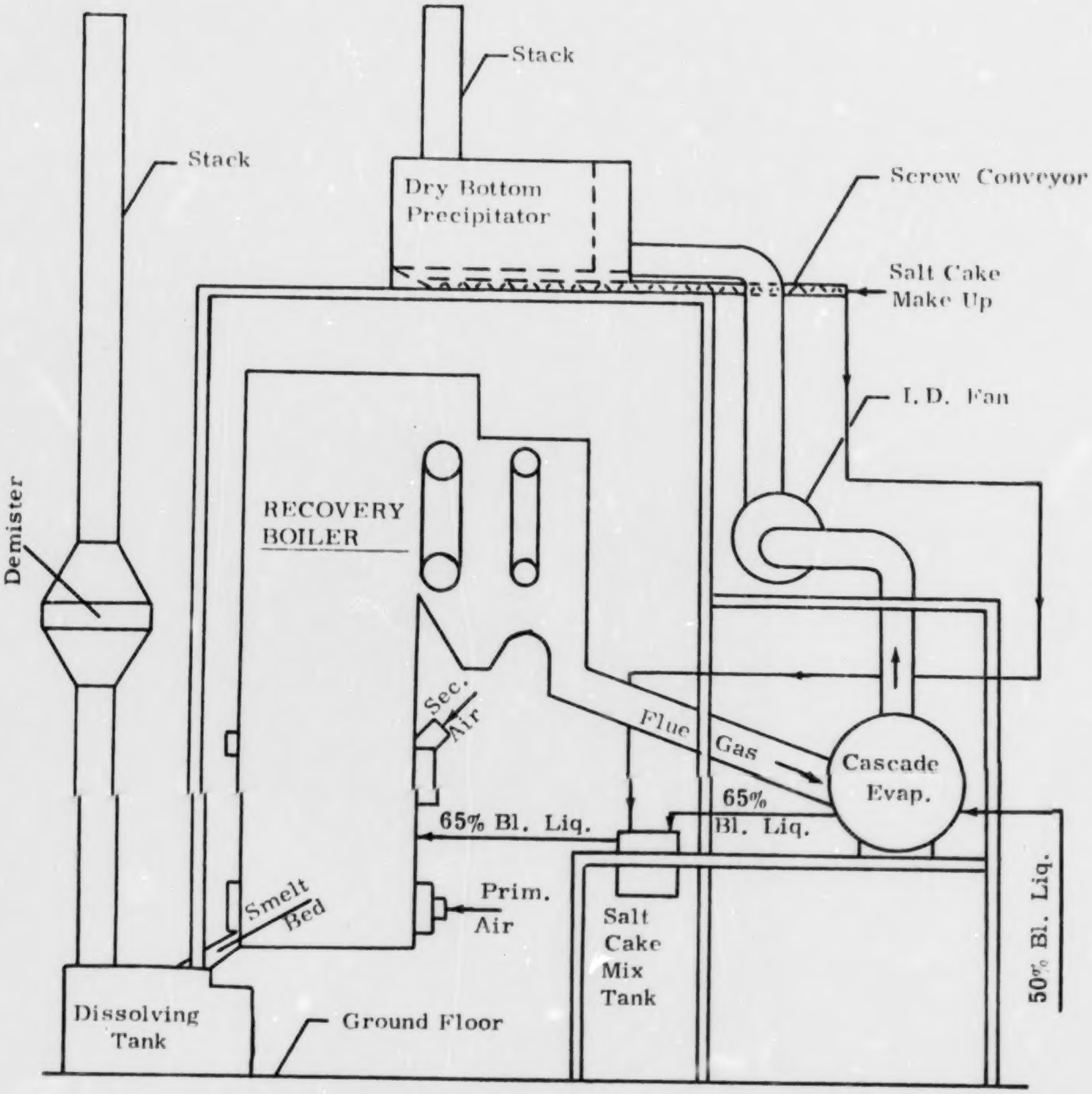
Stack
Stack
Dry Bottom Precipitator
Screw Conveyor
Salt Cake Make Up
I. D. Fan
Demister
RECOVERY BOILER
Sec. Air
Flue Gas
Cascade Evap.
65% Bl. Liq.
65% Bl. Liq.
Prim. Air
Smelt Bed
Salt Cake Mix Tank
50% Bl. Liq.
Dissolving Tank
Ground Floor

Figure 16.11. Typical Dry Bottom Electrostatic Precipitator Recovery Boiler Installation.

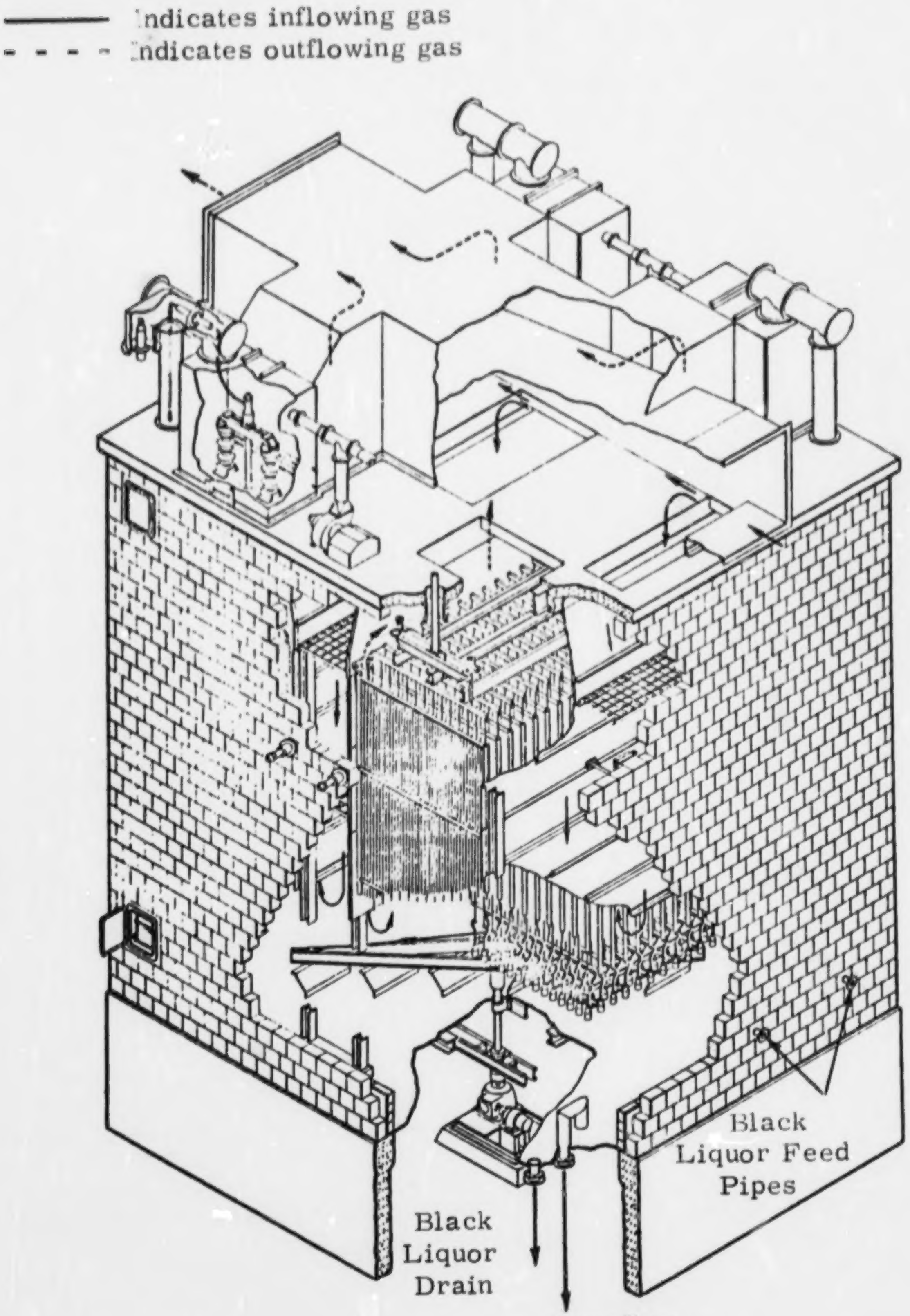
——— Indicates inflowing gas
- - - - Indicates outflowing gas
Black Liquor Feed Pipes
Black Liquor Drain
Overflow

Figure 16.12. Vertical Flow Salt Cake Precipitator with Wet Bottom Dust Removal System.

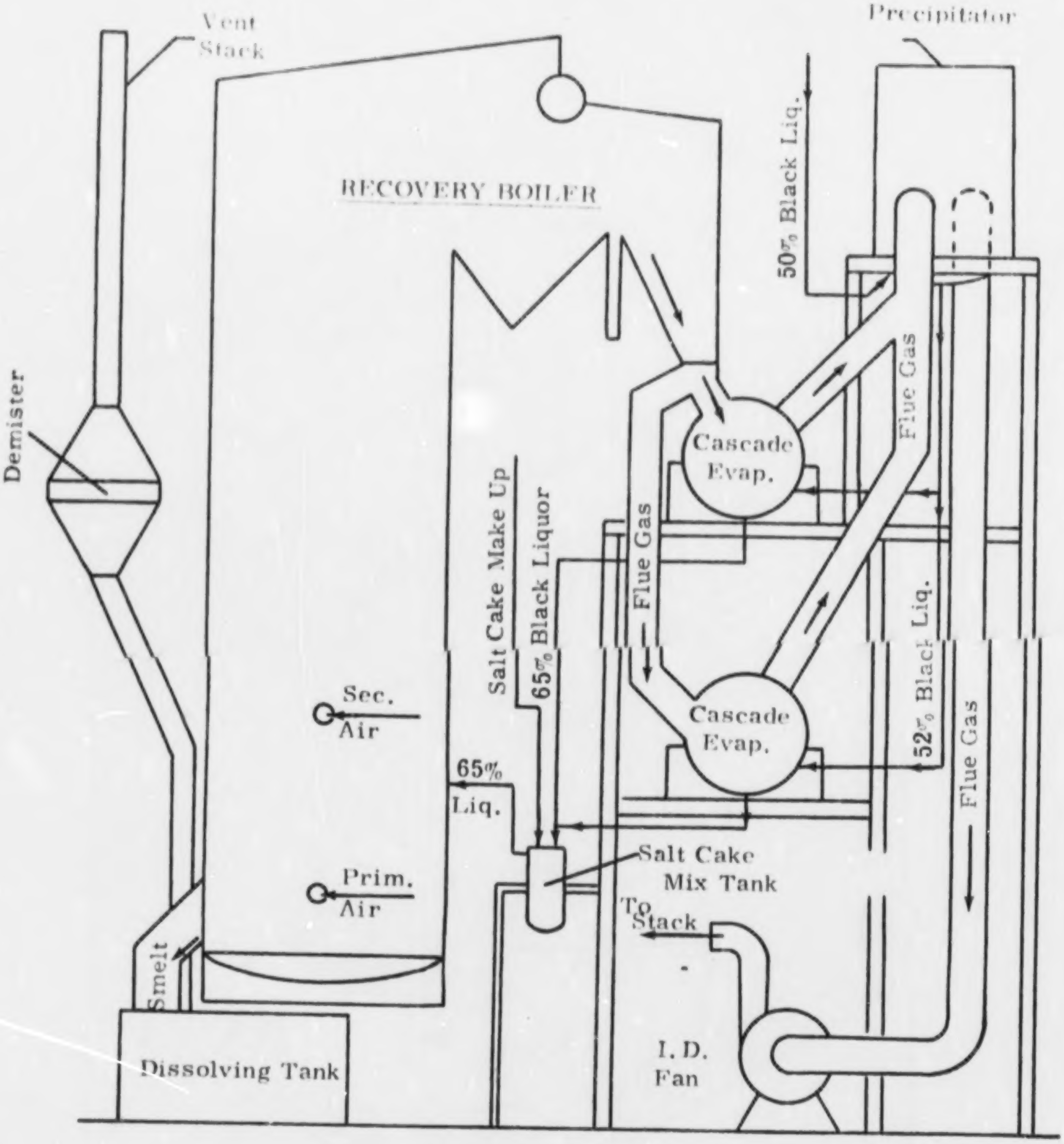
Vent Stack
Precipitator
50% Black Liq.
RECOVERY BOILER
Demister
Flue Gas
Cascade Evap.
Salt Cake Make Up
65% Black Liquor
Flue Gas
Sec. Air
Cascade Evap.
52% Black Liq.
Flue Gas
65% Liq.
Salt Cake Mix Tank
Prim. Air
To Stack
Smelt
I. D. Fan
Dissolving Tank

Figure 16.13. Typical Wet Bottom Electrostatic Precipitator Recovery Boiler Installation.

of salt cake. The wet bottom system takes advantage of the fact that the salt cake is more readily soluble in less concentrated (50%) liquor. This solution is pumped into the collection hopper where the electrode rappings fall, thus facilitating collection, removal, and redissolving of the recovered dust.

Dry bottom precipitators have been primarily of the vertical flow type with general construction features shown in Figures 16.10 and 16.12.

The precipitator shell walls were constructed of hollow, vitrified, glazed tile, with the voids filled with concrete and reinforcing rods to give a solid masonry wall of 8" thickness. The roof of the precipitator was an 8" thick, concrete slab, suitably reinforced. Either structural steel or masonry walls were used to support the entire shell. The collecting electrodes were of the rod curtain type, each consisting of a series of vertical parallel rods, 3/8" diameter, and spaced on 1-1/2" centers. These rods were supported in a pipe framework, suspended in the precipitator at equal spacing to form ducts or gas passages 10" wide through which the gas passed. In the center of each of these ducts was suspended a series of discharge electrodes, held taut by weights at the bottom, and the entire system of discharge electrodes [illegible] spaced at the bottom by [illegible] steadying framework. These discharge electrodes were suspended at the top from an insulated high tension supporting framework.

Both the collecting and discharge systems were equipped with air operated vibrators for removing the collected dust from the electrodes. The hoppers at the bottom were usually constructed of steel which was heat insulated and steam traced to prevent moisture condensation with subsequent cake plugging problems.

Wet bottom precipitators. To alleviate the plugging problem, the wet bottom precipitator was conceived and is now used throughout the pulp and paper industry. In this system, the 50% liquor from the multiple-effect evaporator is pumped into the bottom of the precipitator where the collected particles fall and are dissolved as shown in Figure 16.13.

Present day precipitator design utilizes a horizontal flow type unit as shown in Figure 16.14. This arrangement provides more electrical sectionalization and longer treatment time for greater efficiency. Shell construction is of tile, concrete, or specially insulated and painted steel.

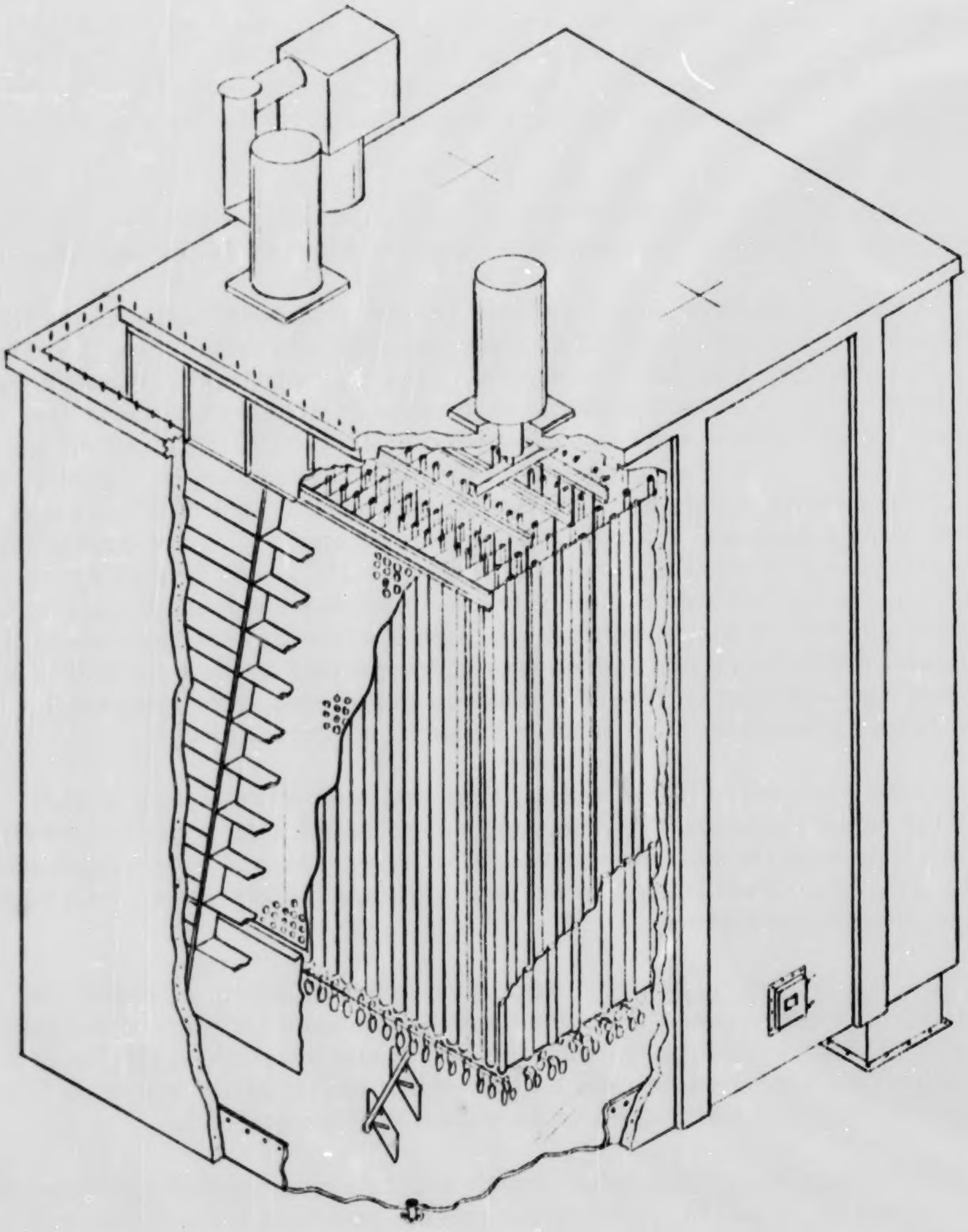

Figure 16.14. Horizontal Flow Salt Cake Precipitator with Wet Bottom Dust Removal System.

Several recovery boilers have been installed without an electrostatic precipitator. These systems use high energy venturi scrubbers as shown on Figure 16.15. The 50% liquor from the multiple-effect evaporator is pumped into a venturi scrubber on the main flue gas line from the boiler where particulate matter from the flue gas is scrubbed out and dissolved in the liquor. The stream is then routed to a cyclone separator where the gas is separated and sent to a stack, and concentrated liquor is pumped to the salt cake mix tank.

The principal disadvantage in this system is the high power required due to the high pressure drop on the flue gas. The emission of saturated gas from the stack can also be a problem.

Figure 16.16 shows a system utilizing both a cyclone evaporator and an electrostatic precipitator. In this system, the flue gas passes through a cyclone evaporator where it is scrubbed by the 50% liquor from the evaporators in a tangential inlet spray system. The flue gas then goes to an electrostatic precipitator for further particulate recovery. The precipitator is generally a wet bottom type and uses 50% liquor from the multiple-effect evaporator for collecting the rappings. The liquor is concentrated to approximately 65% solids in the cyclone evaporator by the action of the hot flue gas. It is then pumped to the salt cake mix tank for feeding to the boiler.

Contact of the black liquor with the effluent gases appears to be a major contributing factor in the relase of odors from paper mills. Because of the emphasis on odor control, recent recovery boiler designs do not utilize direct contact evaporators. Instead, the flue gas is cooled in air preheaters or enlarged economizer sections in the boiler.

Factors affecting use of electrostatic precipitators in the pulp and paper industry. The type of wood cooked, the yield, and the type of paper to be made from the pulp all affect the cooking procedure in the kraft pulp mill. These factors also affect the amount of emissions from the recovery boiler, which, in turn, affects the use of a precipitator.

A large portion of the pulp made today is bleached, which requires that the pulp be cooked to a lower yield. In turn, this requires more cooking liquor and more recovery of chemicals. Therefore, a mill making all bleached kraft requires a larger precipitator for the same tonnage of pulp than an unbleached kraft mill.

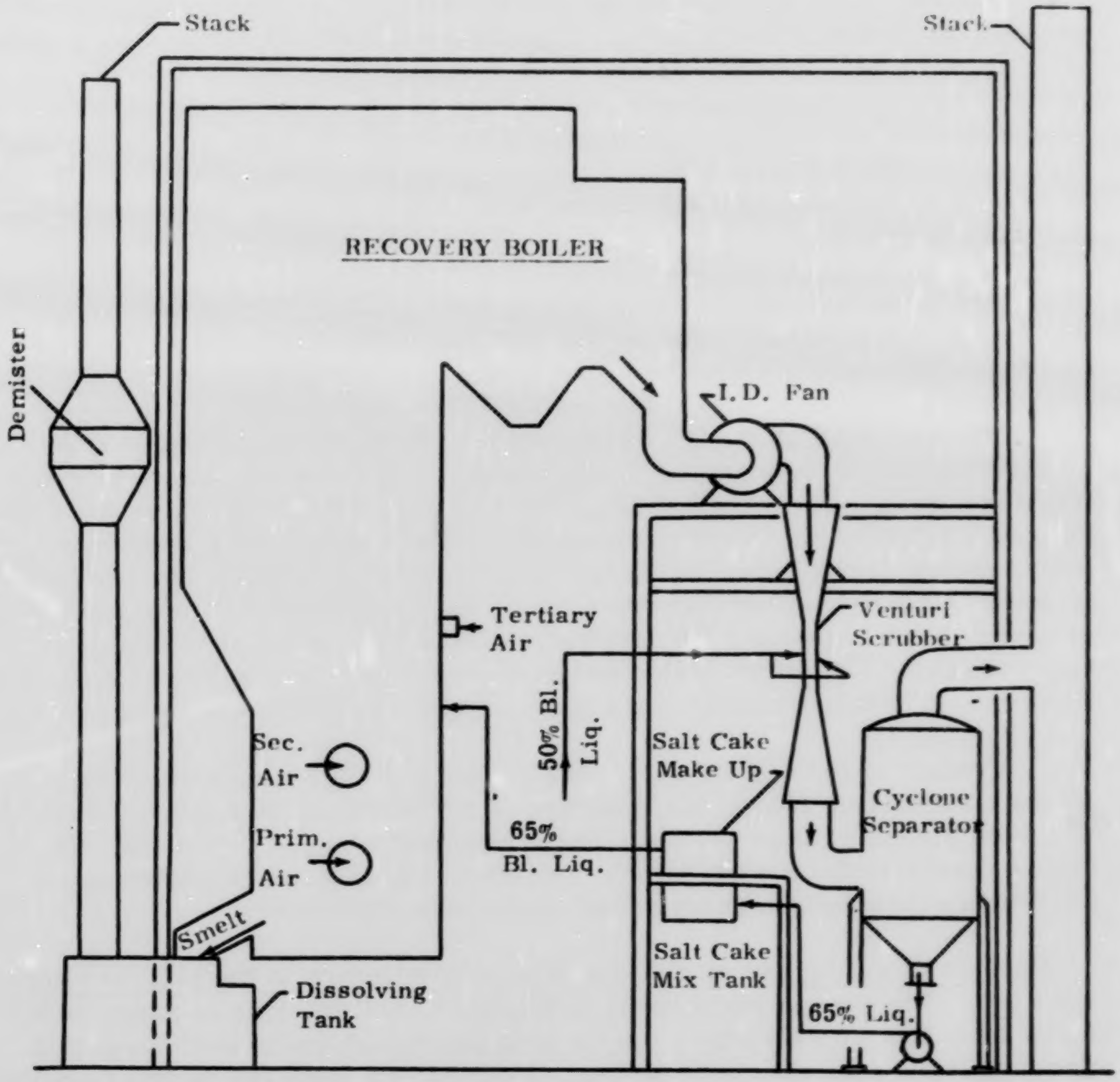
Stack
Stack
RECOVERY BOILER
Demister
I. D. Fan
Tertiary Air
Venturi Scrubber
50% Bl. Liq.
Salt Cake Make Up
Sec. Air
Prim. Air
Cyclone Separator
65% Bl. Liq.
Smelt
Salt Cake Mix Tank
Dissolving Tank
65% Liq.

Figure 16. 15. Typical Venturi Scrubber Recovery System Installation.

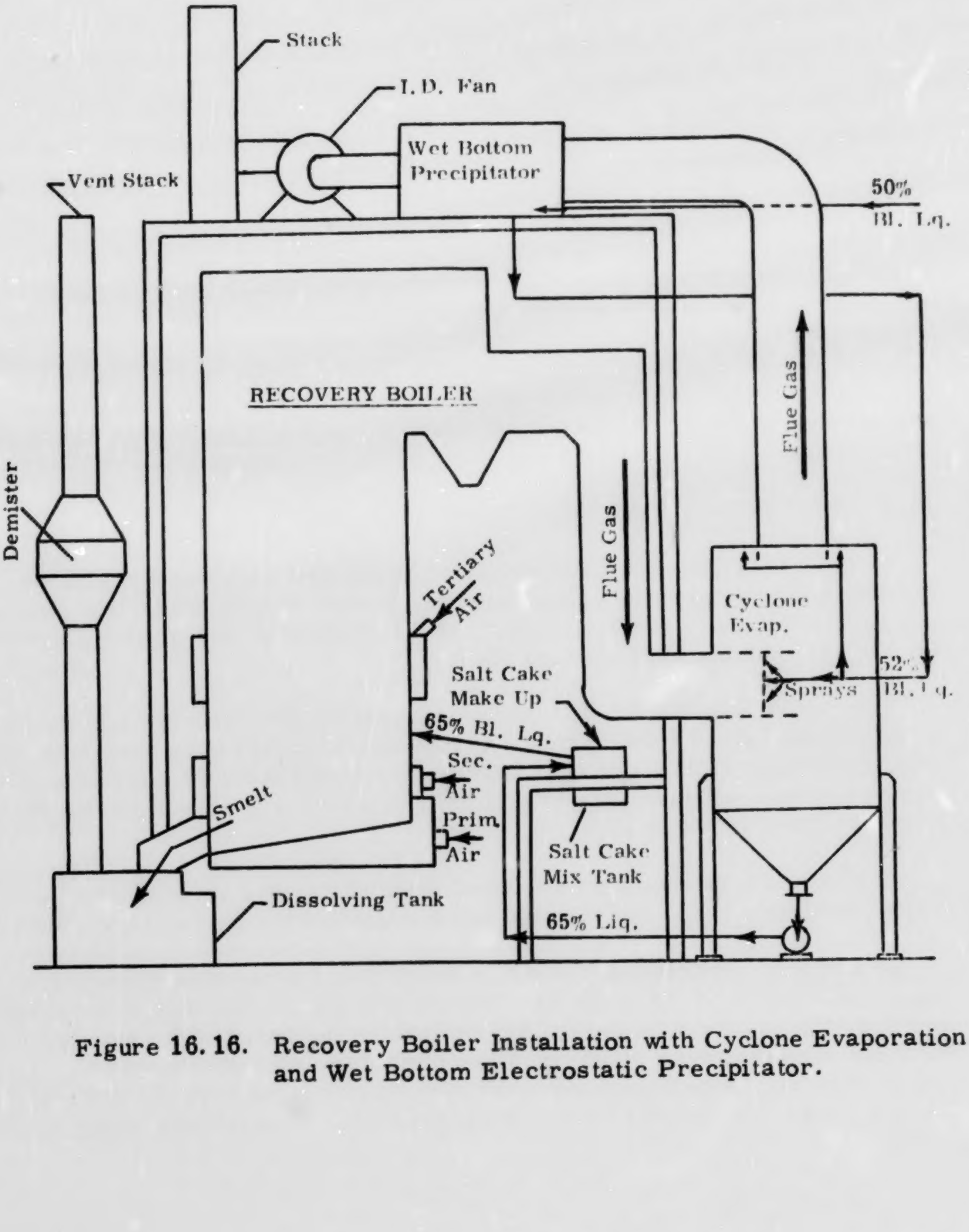
Stack
I. D. Fan
Wet Bottom Precipitator
Vent Stack
50%
Bl. Lq.
Demister
RECOVERY BOILER
Flue Gas
Flue Gas
Tertiary Air
Cyclone Evap.
52%
Bl. Lq.
Sprays
Salt Cake Make Up
65% Bl. Lq.
Sec. Air
Prim. Air
Smelt
Salt Cake Mix Tank
Dissolving Tank
65% Liq.

Figure 16.16. Recovery Boiler Installation with Cyclone Evaporation and Wet Bottom Electrostatic Precipitator.

The main types of unbleached kraft produced are bag paper and linerboard for the manufacture of cardboard boxes. The bag paper requires more cooking than the linerboard from which the highest yield of all the kraft paper is obtained. Therefore, a bag paper plant would require more recovery of chemicals and larger precipitators than a plant producing linerboard exclusively.

The semichemical mills, if they do not produce kraft paper in the same mill, would need very little recovery of chemicals and may (if ammonia base cooking is used) eliminate the need for a precipitator, since a very small amount of particulate matter is emitted.

Some of the larger sulfite mills, especially those using magnesium base chemicals, are now installing chemical recovery boilers and must use precipitators.

16.5 PRECIPITATOR APPLICATIONS

The use of electrostatic precipitation for salt cake recovery in the United States has generally paralleled the growth of pulp production utilizing the sulfate or kraft process. Figure 16.17 illustrates this growth pattern for the period from 1939 through 1969.

It is estimated that a total of approximately 278 recovery boilers has been installed since 1936. Of those, approximately 181 have been equipped with electrostatic precipitators, handling a total of about 35 million cu ft of gas per minute. A summary of pulp mill recovery systems in the United States is given in Table 16.4.

Figure 16.18 shows the installed precipitator gas volume over the 1939-1969 period, and the average volume handled for 5-year increments.

Economics. Precipitator costs on an FOB and erected cost basis are shown in Figures 16.19 to 16.21. The FOB costs are the flange-to-flange precipitator costs corrected to 1969 dollars on the basis of the average wholesale price index. FOB costs are plotted for all installations for which data were available, and cover efficiency ranges from 90 to 99+%. Erected cost data include cost of the precipitator, foundations, and erection.

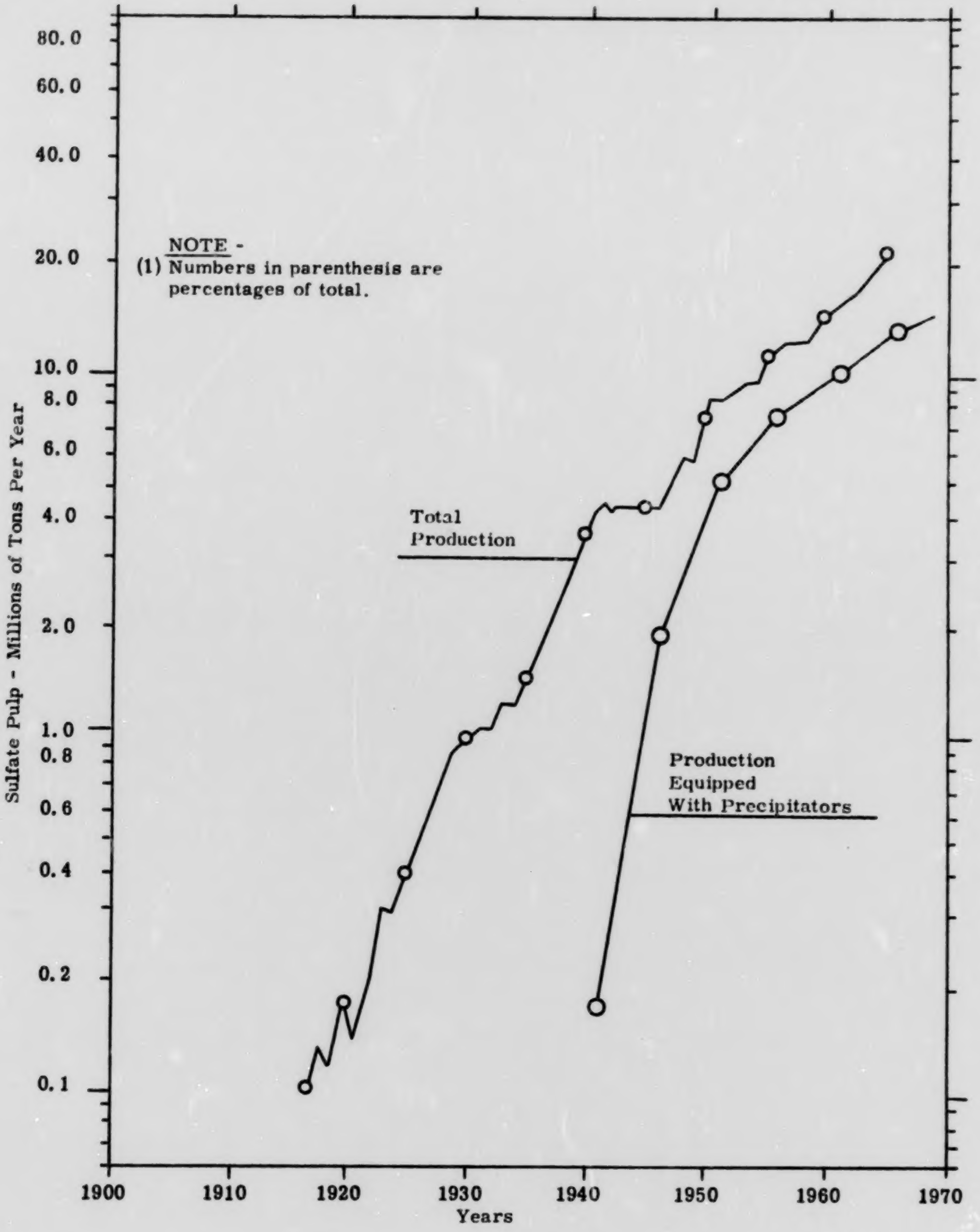

Figure 16.17. Estimate of Total Sulfate Pulp Production with Electrostatic Precipitators Installed on Recovery Furnace Gases.

TABLE 16.4

SUMMARY OF PULP MILL RECOVERY SYSTEMS

Region	Number of Recovery Boilers					Precipitator Installed		Precipitator Bottom		Evaporator Type		
	SO_4	Acid SO_3	Soda (NaOH)	Semi-Chem.	MgO	Yes	No	Wet	Dry	Cascade	Cyclone	Venturi
NORTHEAST	22					13	9	5		3	1	5
		2	2	2	2	3	2	3		-	1	2
Maine, Mass., N. H., N. J., N. Y., Pa., Md.												
Totals	22	2	2	2	2	16	11	8		3	2	7
SOUTHEAST	143					108	35	47		30	9	15
			1	6		3	4					
Va., N. C., S. C., Ga., Fla., Ala., Miss., Ky., Tenn.												
Totals	143		1	6		111	39	47		30	9	15
SOUTHWEST	51					28	23	19		14	20	8
				3		3						
La., Ark., Tex., Okla.												
Totals	51			3		31	23	19		14	20	8
MIDWEST	12					9	3	8		1		3
		4		3		2	4	2		2	2	1
Mich., Ohio, Wisc., Minn., Ind., Iowa												
Totals	12	4		3		11	7	10		3	2	4
NORTHWEST	42					20	22	13		6	4	9
		2		1	10	1	12			2	2	
Wash., Ore., Idaho, Mont., Alaska												
Totals	42	2		1	10	21	34	13		8	6	9
FAR WEST	8					3	5				4	2
Calif., Ariz.												
Totals	8					3	5				4	2
GRAND TOTALS	278	8	3	15	12	193	119	97		58	43	45

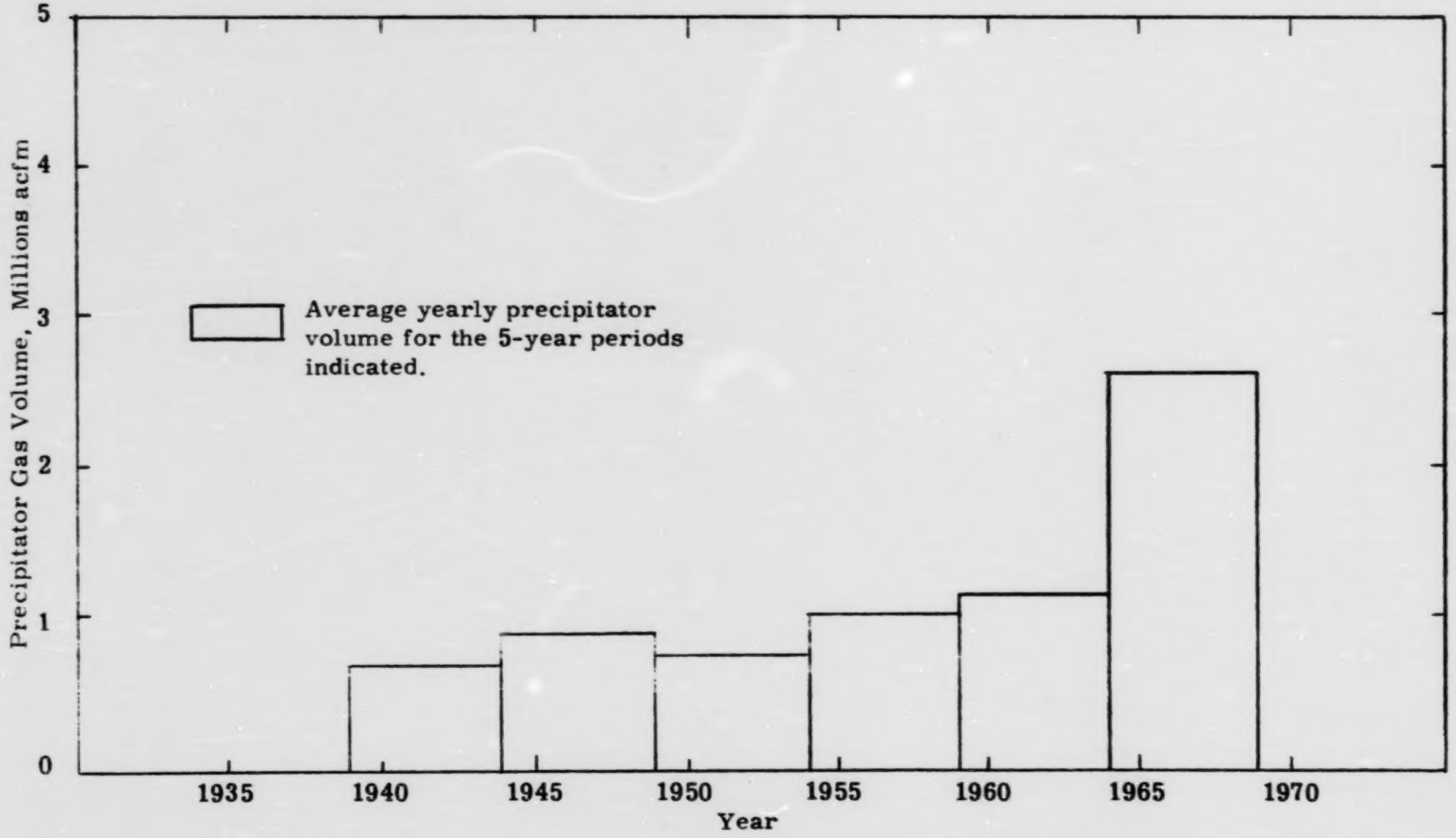

Figure 16.18. Installed Precipitator Gas Volume Over the Period 1939 through 1969.

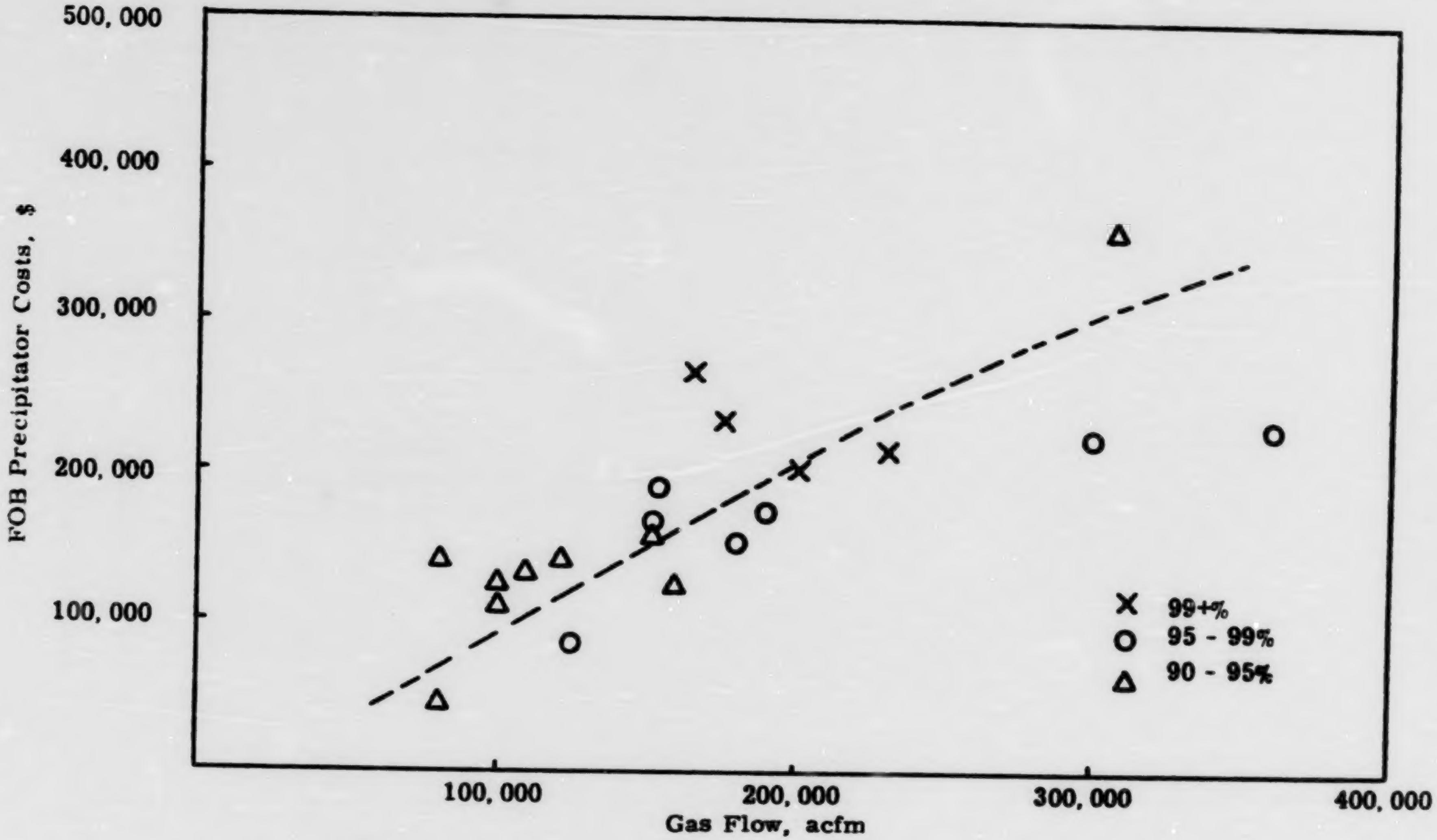

Figure 16.19. FOB Costs for Recovery Boiler Precipitator (1969 Dollars).

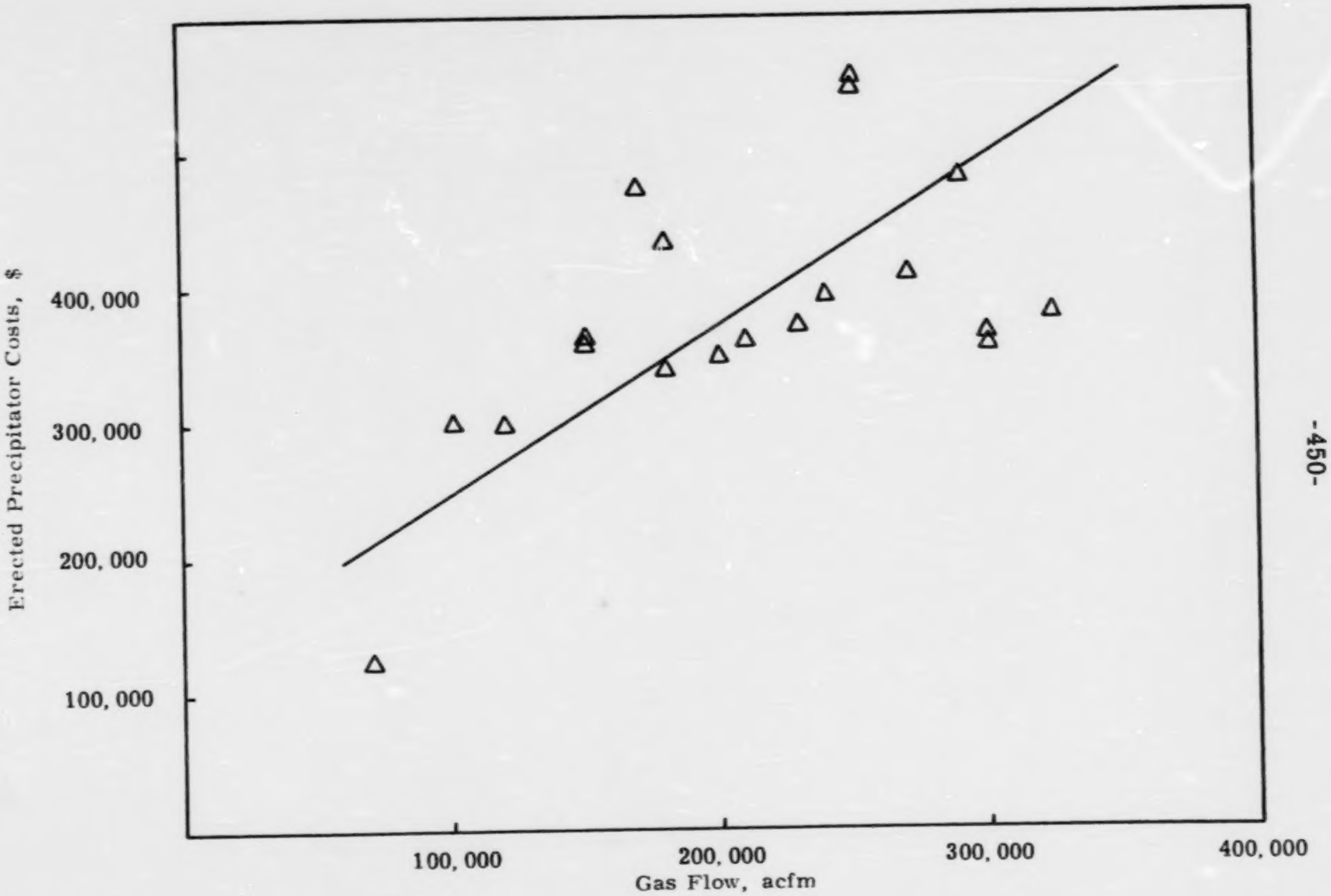

Figure 16.20. Erected Costs of Recovery Boiler Precipitators (1969 Dollars - 99+% Efficiency).

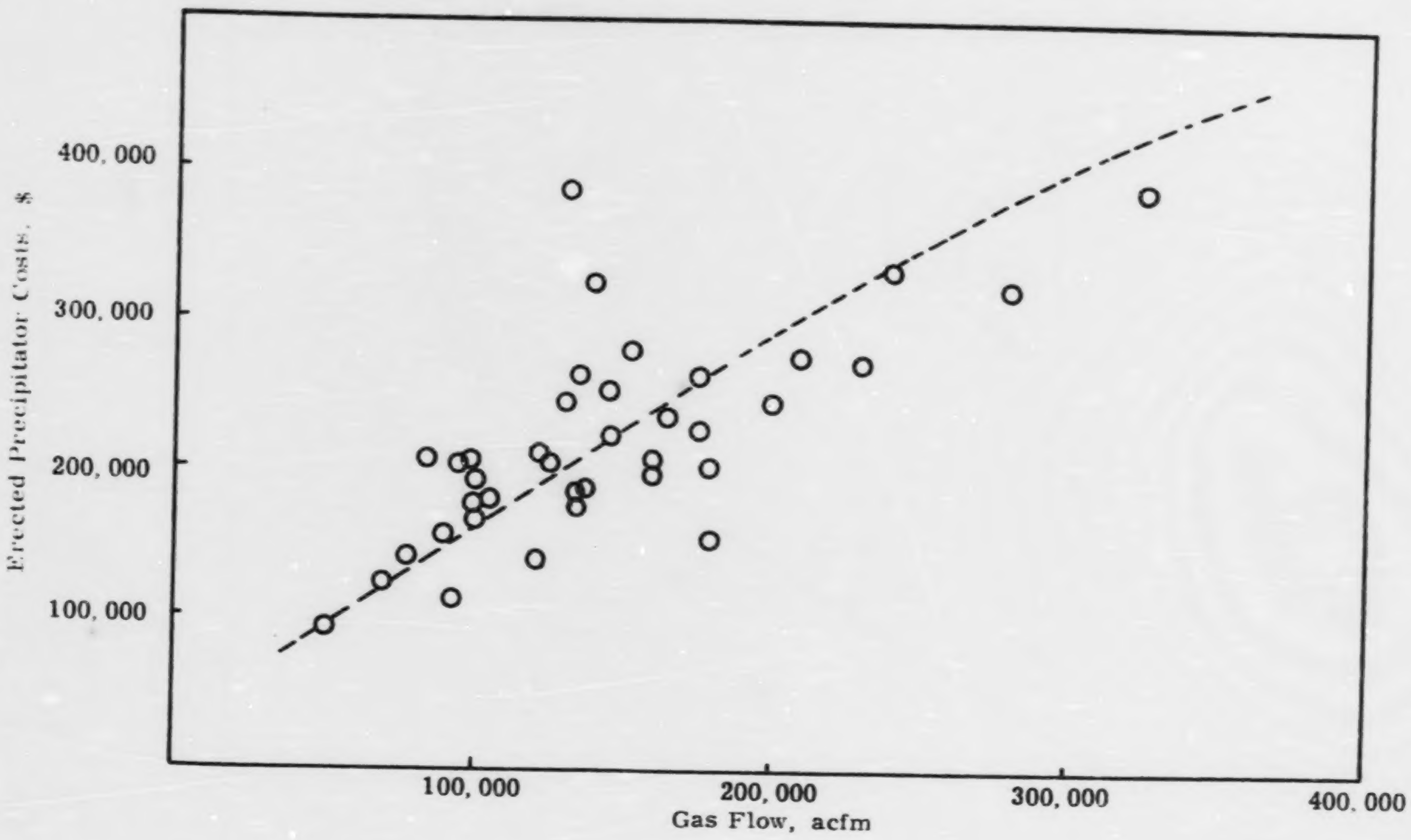

Figure 16.21. Erected Costs for Recovery Boiler Precipitators (1969 Dollars - 90-99% Efficiency).

FOB cost data indicate the increase in cost for precipitators designed for 99% efficiency over that for lower efficiency units. However, sufficient data are not available to statistically differentiate the higher costs associated with the higher efficiency precipitators. Since more information is available on erected costs, the higher costs for precipitators designed for greater than 99% efficiency are apparent. In the 90-95 and 95-99% range, variations in costs as a result of other factors mask the effect of efficiency variations. Consequently, the data are plotted on the same curve with the efficiency range noted.

Variations in costs of precipitators for recovery boilers can be attributed to a number of factors, the most important ones being:

(1) The size of the precipitator as related to dust characteristics (resistivity, particle size) and gas conditions (temperature, which indirectly affects dust resistivity, moisture content);

(2) The geographical location of the installation particularly for erected cost, since labor costs vary greatly throughout the country;

(3) The pricing-profit policy of the various competitors bidding on the job;

(4) Whether the installation is a backfit, rebuild, or upgrading of an existing installation, which may require additional engineering costs to "shoehorn" the precipitator into the overall installation;

(5) Whether the installation is ground level, above ground level, indoors, outdoors, etc.; and

(6) Whether the precipitator is a wet or dry bottom, stainless steel or tile shell.

The overall average precipitator costs (where data are available) for the two 5-year increments from 1960 through 1969 have been summarized in Table 16.5 and graphically shown in Figure 16.22. The majority of the installations are contracted for on an erected basis, thus the lack of FOB costs.

Operating costs. Practice varies between plants on the labor allocated to the maintenance of electrostatic precipitators in pulp and paper

Table 16.5

Electrostatic Precipitator Costs (1960 through 1969)

Year	Gas Volume Range in 1000's acfm	Design Eff. Prorated on acfm Basis - %	Cost in cents/acfm	
			Erected	FOB
1960	60-160	94.8	156 (7)	No data
1961	70-300	97.7	121 (3)	62 (1) (a)
1962	78-211	95.9	150 (6)	46 (2)
1963	49-180	94.2	137 (6)	72 (1)
1964	85-360	97.0	171 (7)	69 (4)
1965	90-305	97.2	125 (9)	No data
1966	120-868	98.2	113 (9)	86 (3)
1967	105-263	98.7	150 (4)	No data
1968	105-355	99.0	166 (13)	163 (1)
1969	120-325	99.5	186 (7)	No data
Overall Averages	88-333	97.7	149	77*
Average 1960-1964	68-242	96.2	151	63*
Average 1965-1969	108-423	98.4	148	105*

Note: (a) Numbers in parenthesis are sample size, i.e. number of installation contract prices averaged to obtain indicated costs/acfm (all precipitator manufacturers represented).

* Based on incomplete data above.

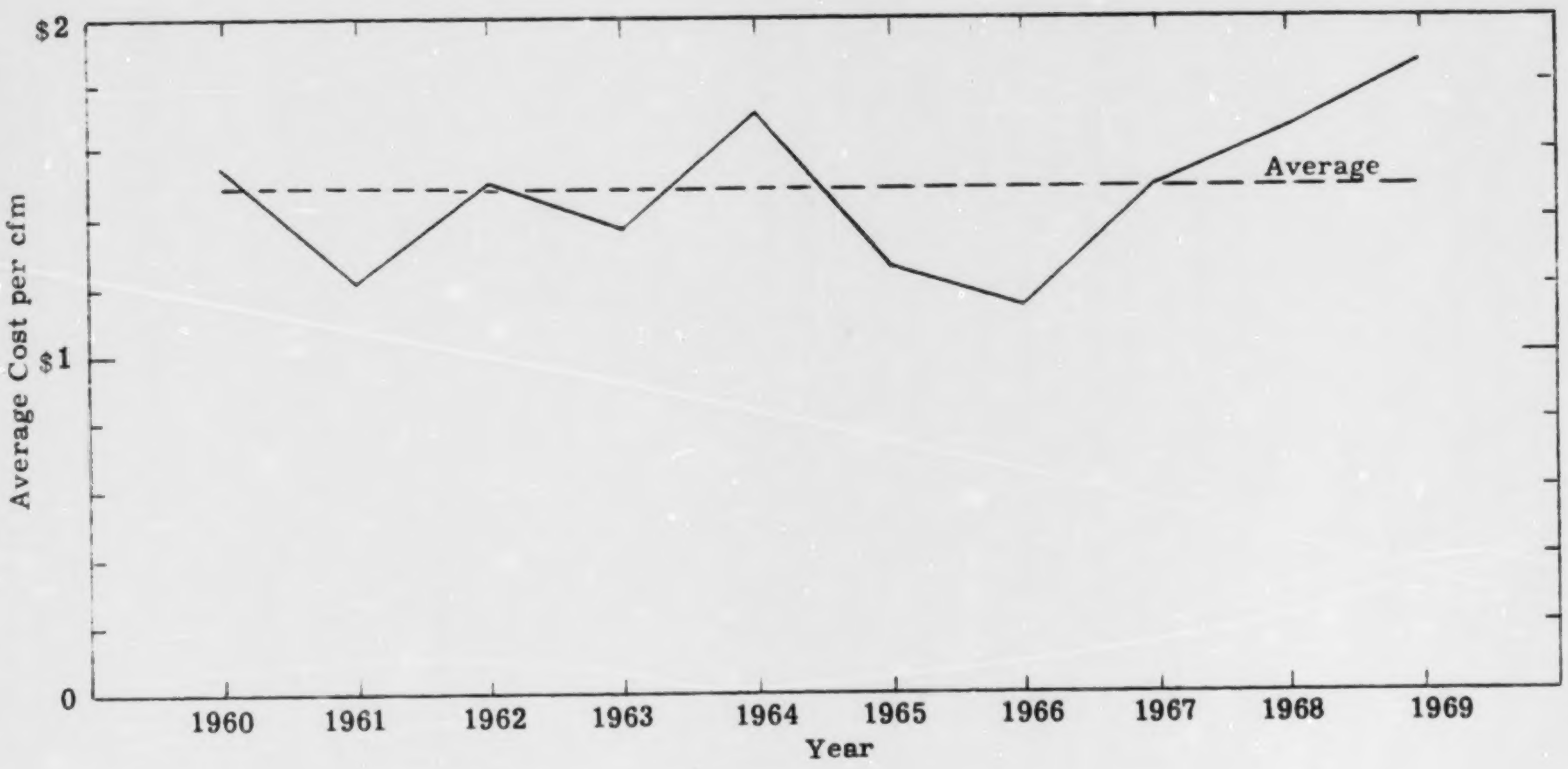

Figure 16.22. Electrostatic Precipitator Costs (1960-1969).

mills. In general, precipitators are cleaned and inspected during normal plant shutdowns, which often correspond to national holidays, which means that there are 6-10 maintenance periods per year. In addition to these scheduled periods, removal of broken wires and adjustment and servicing of rappers require maintenance labor.

Precipitators operate unattended for the most part, and operating labor is not generally charged against the precipitator costs, although log keeping and periodic adjustment of controls do require some time.

Table 16.5 shows the total operating costs computed for two recovery boiler precipitators handling gas flows of 250,000 and 100,000 cfm. Maintenance costs were taken from records of two plants and constitute around 1,000 man hours per year for the 250,000 cfm unit, and 650 man hours for the 100,000 cfm precipitator, assuming $6.00 per man hour labor costs.

Power consumption was based on the power used, as computed from the data shown in Figure 16.23. Computation was based on 7,000 hours operation per year and $0.005/kWh energy costs. The energy costs do not include draft fan power. Capital costs were based on 12.8% of the total precipitator investment, which was computed on the basis of the data given in Figure 16.20 (99% erected cost curve).

	Case 1	Case 2
Gas flow, cfm	250,000	100,000
Efficiency, %	99	99
Maintenance costs, $	5,929	3,850
Precipitator power used, kW	150	60
Precipitator power costs, $	5,250	2,120
Operating labor	--	--
Capital costs, $	56,400	32,000
Total Annual Costs, $	67,579	37,970

16.6 SUMMARY OF PRECIPITATOR OPERATING CONDITIONS

Operating conditions from electrostatic precipitator installations on recovery furnaces in the sulfate pulp industry have been summarized in a series of histograms. Figures 16.24 through 16.28 represent the frequency

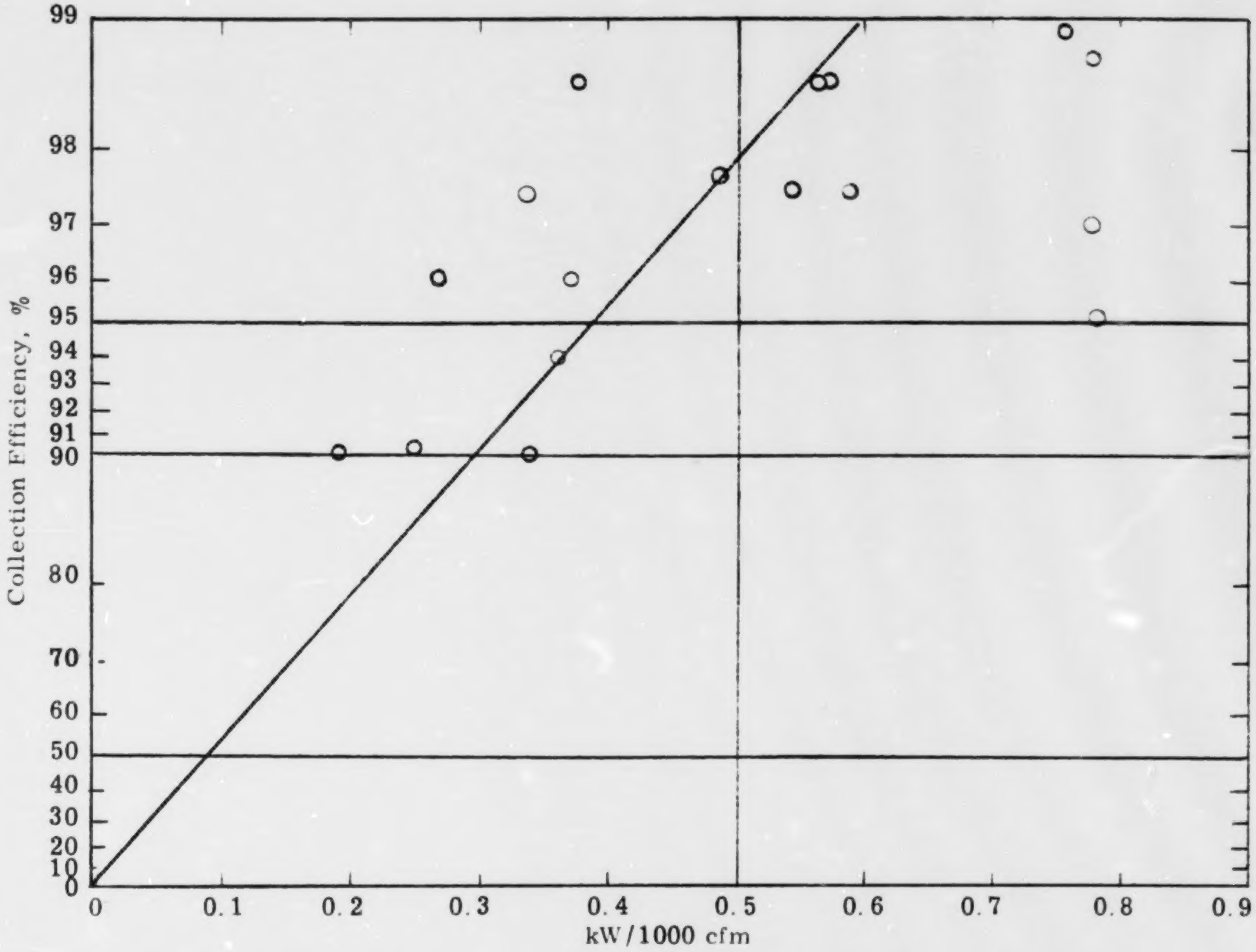

Figure 16.23. Collection Efficiency Vs. Input Power per Thousand cfm in Pulp and Paper Installations.

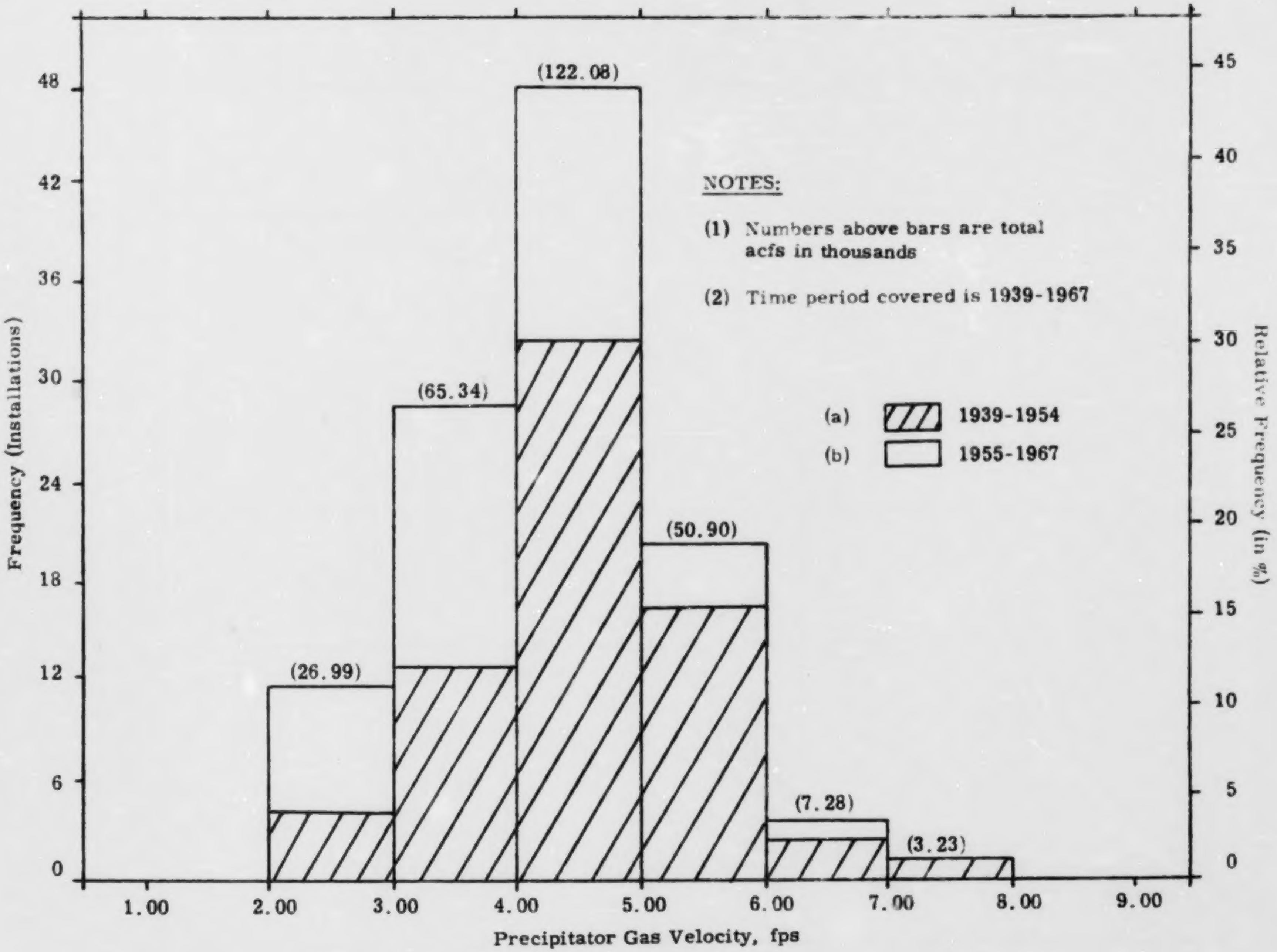

Figure 16.24. Distribution of Precipitator Gas Velocity.

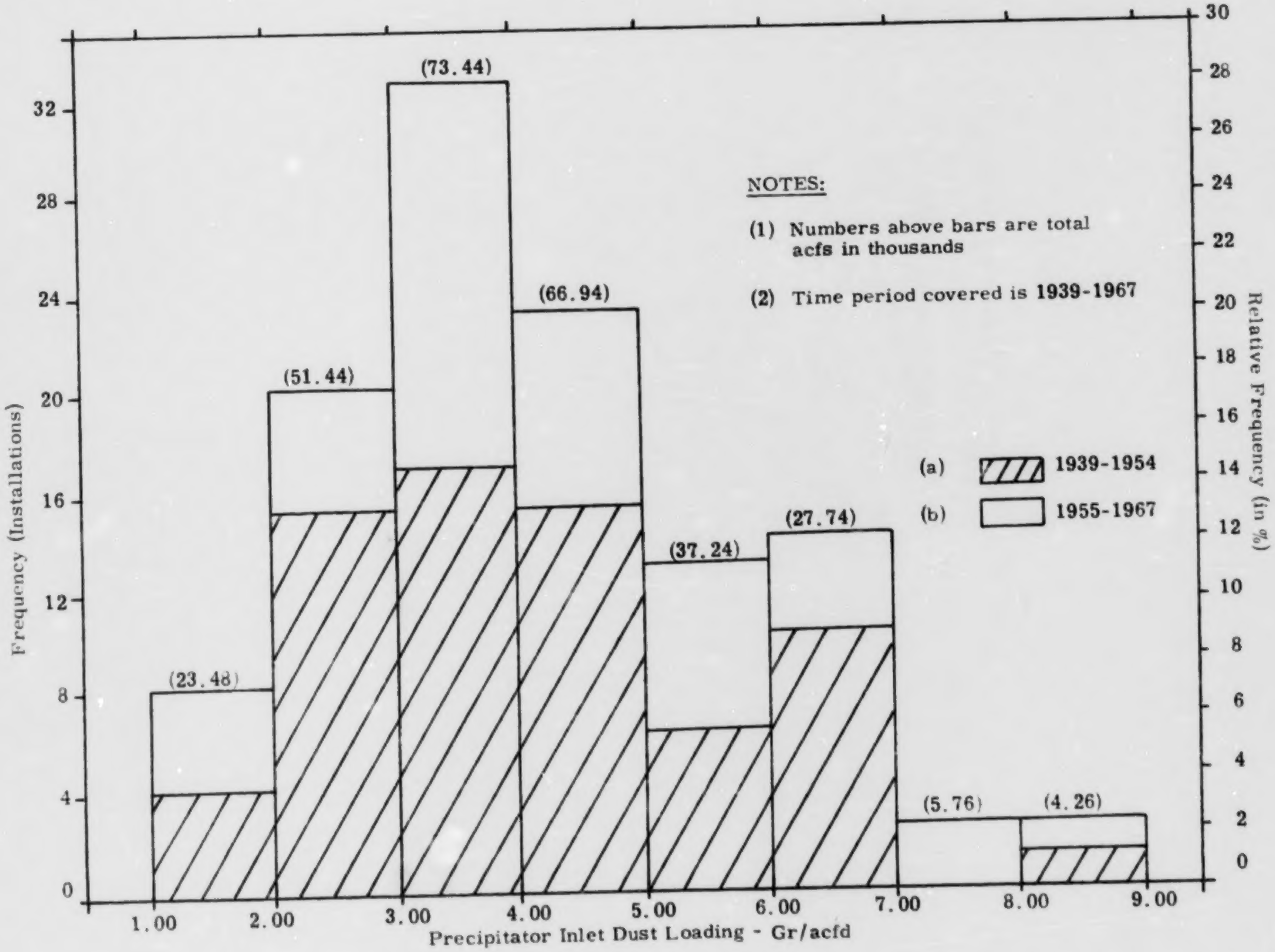

Figure 16.25. Distribution of Precipitator Inlet Dust Loading.

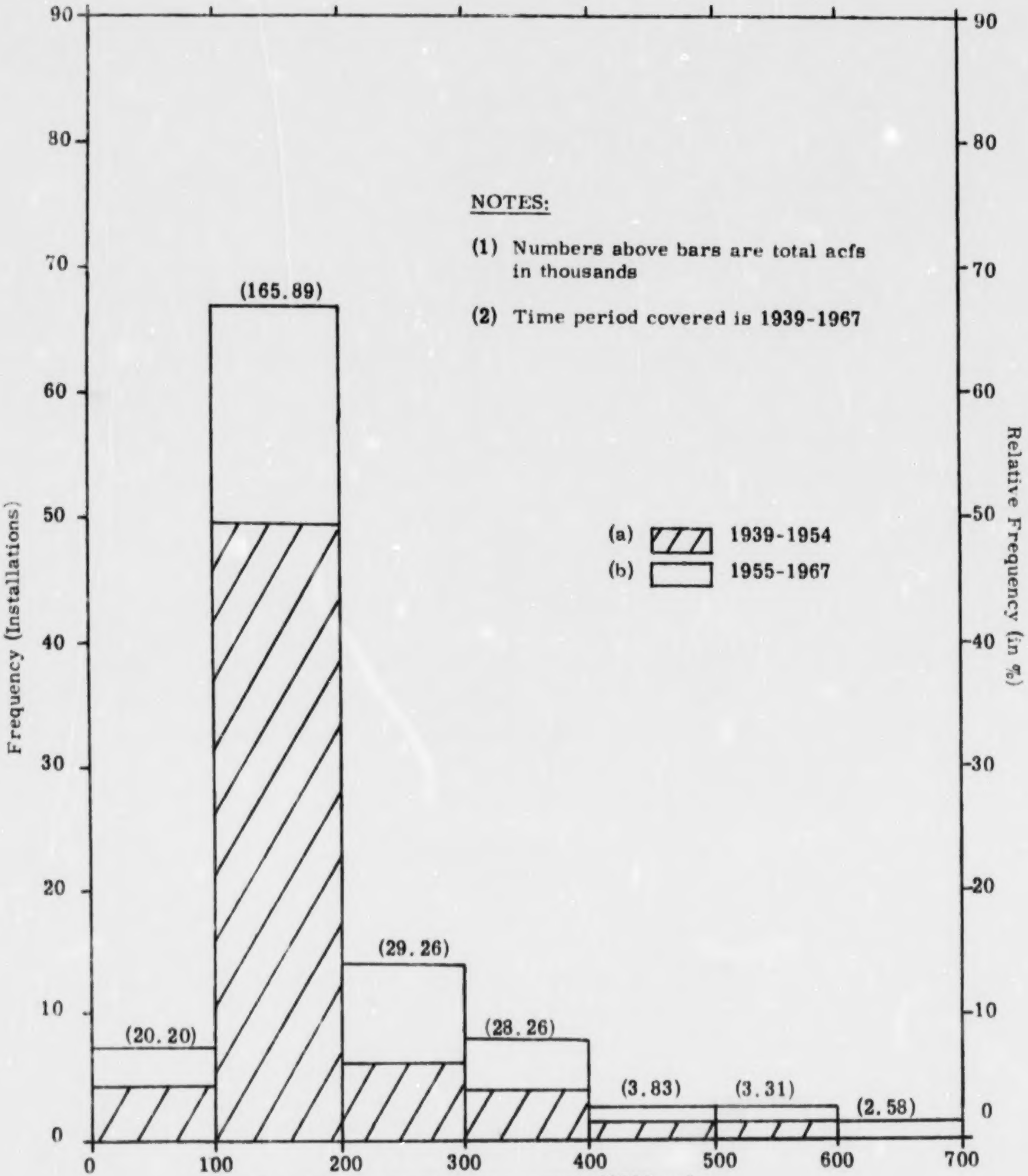

Figure 16.26. Distribution of Precipitator Connected Input Power.

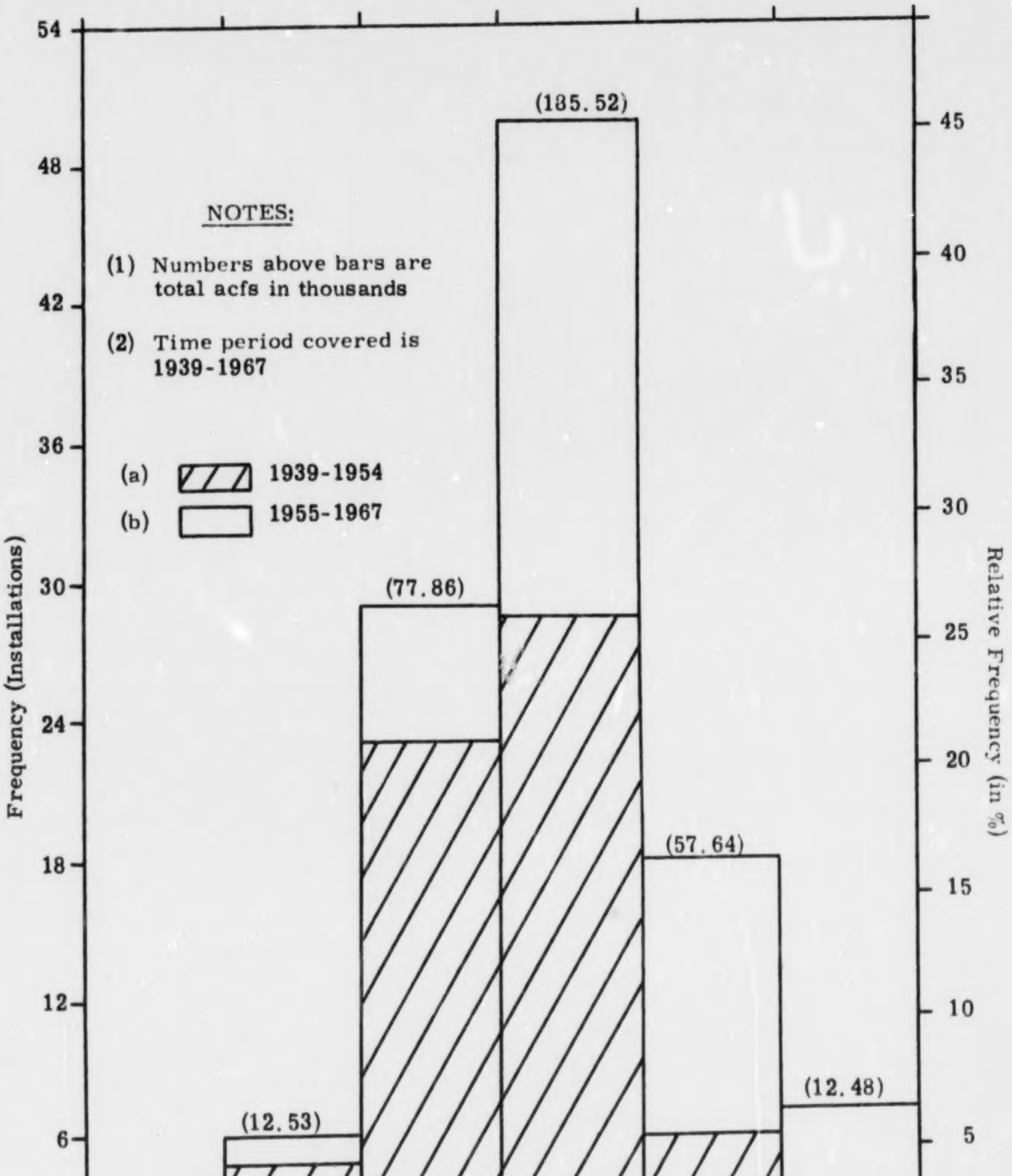

Figure 16. 27. Distribution of Precipitator Inlet Gas Temperature.

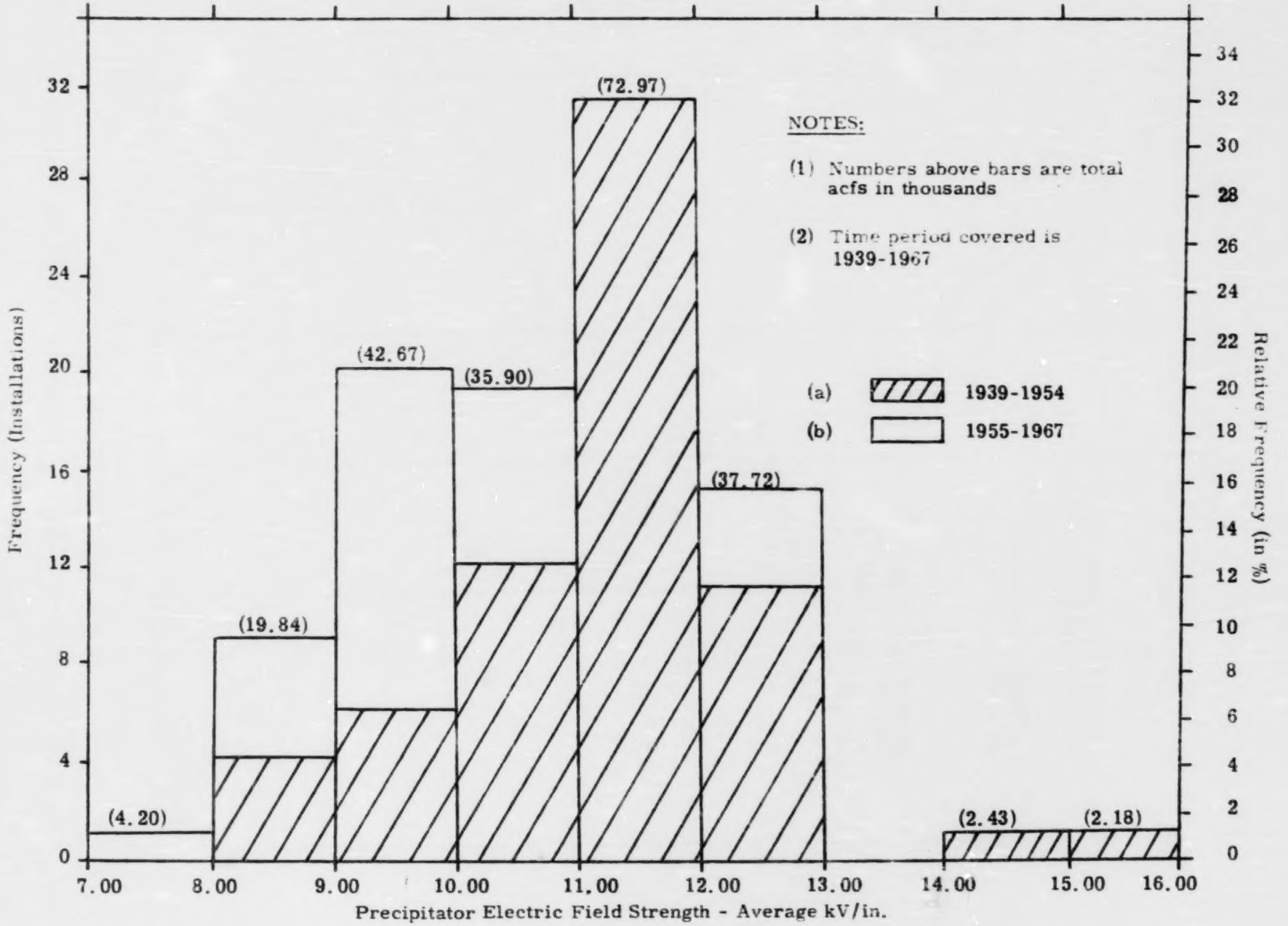

Figure 16.28. Distribution of Precipitator Design Electrical Field Strength.

of appearance of the particular parameters indicated. Included in the five parameters are; gas velocity, inlet dust loading, input power, gas temperature, and electric field strength.

The data have been divided into two time periods, 1939 through 1954 and 1955 through 1967. In addition, the gas volume flow handled within each interval has been included.

16.7 ASSESSMENT OF THE STATE OF THE ART

Resume of contemporary practical design methodology. The following is a resume of practical design methodology for selecting electrostatic precipitation equipment to remove salt cake from recovery furnace gas effluent. There are five basic technical considerations in specifying precipitation equipment for this application, namely: customer specifications, precipitator type and size, electrical energization equipment, type and quantity of electrode rappers, and precipitator dust removal system.

Ideally, design methodology should result in specifying equipment that first satisfies the requirements of the customer, and second, is either within economic limitations of sound competitive bidding or can be economically justified for noncompetitive situations. Nearly all proposed salt cake collection equipment at this time involves both competitive bidding and guaranteed performance requirements, which essentially forces equipment manufacturers to continually upgrade their design methodology to more accurately predict the performance of specified equipment. Due to the competitive situation, equipment manufacturers consider the details of design methodology as proprietary information, each having their own repertoire of design factors and detailed methodology. Thus, the resume presented is general in nature including basic parameters believed to be common to all equipment manufacturers and omitting proprietary detail which varies among manufacturers. The resume is also limited to the precipitator only and does not cover other items such as flues, support steel, etc., many times included in a purchase order.

Customer requirements. Customer requirements are generally presented as formal or informal specifications. Typically, formal specifications state the problem to be solved as well as any special design requirements. Key items expected from the customer include: type and number of boilers, gas volume to be treated, gas temperature, efficiency required, space limitations for equipment, and structural requirements.

In effect, contract operating conditions, extent of equipment to be supplied, and structural requirements are outlined, thus giving sufficient information for equipment manufacturers to propose equipment required to meet a specific problem or application.

Precipitator type and size. Precipitators for salt cake collection are of the horizontal flow, plate type having a casing of rectangular configuration. The design may utilize either insulator compartments or penthouse type construction for enclosing high voltage insulators. Typically, the transformer-rectifiers are mounted on the precipitator roof with control equipment located indoors at a convenient location. The number of bus sections, cells, fields, and chambers of the precipitator are consistent with the design and size requirements for a given application.

The general approach to the design of electrostatic precipitators for the pulp and paper industry follows Method 1 as described in Chapter 9 of Part I on Fundamentals. The selection of the precipitation rate parameter is based on the past experience of the company, but the variation in the value is much smaller than that for electric utilities. A histogram of the precipitation rate parameter for a group of 15 installations is shown in Figure 16.29. The limited range for the value for this parameter is thought to be associated with the very similar particle size distribution, chemical consistency and values of resistivity below the critical range of 2×10^{10} ohm-cm for the recovery boiler in the pulp and paper installation.

The linear relationship between the precipitation rate parameter and power density on the collection electrode that was evident in the electric utilities is not evident from the limited data available for pulp and paper installations. Figure 16.30 shows that this parameter is almost independent of power density for the 15 installations where data were available.

The determination of the collection electrode area is made from the Deutsch-Anderson efficiency equation, $\eta = 1 - \exp(-\frac{A}{v_g} w)$. The efficiency is specified at a particular flow rate; and the precipitation rate parameter has been estimated. The collection electrode area A then computed from solving the efficiency equation

$$A = \frac{v_g}{w} \ln \frac{100}{100 - n} .$$

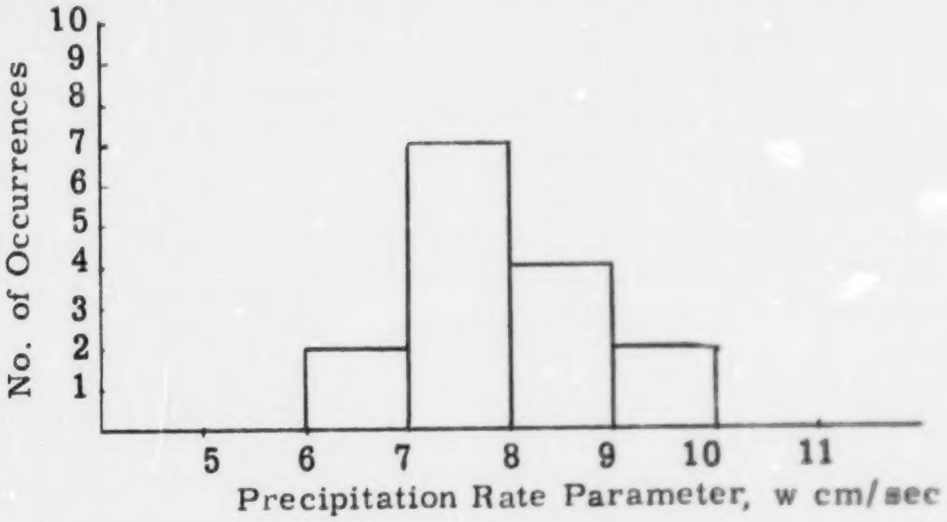

Figure 16.29. Histogram of Precipitation Rate Parameters for 15 Pulp and Paper Installations.

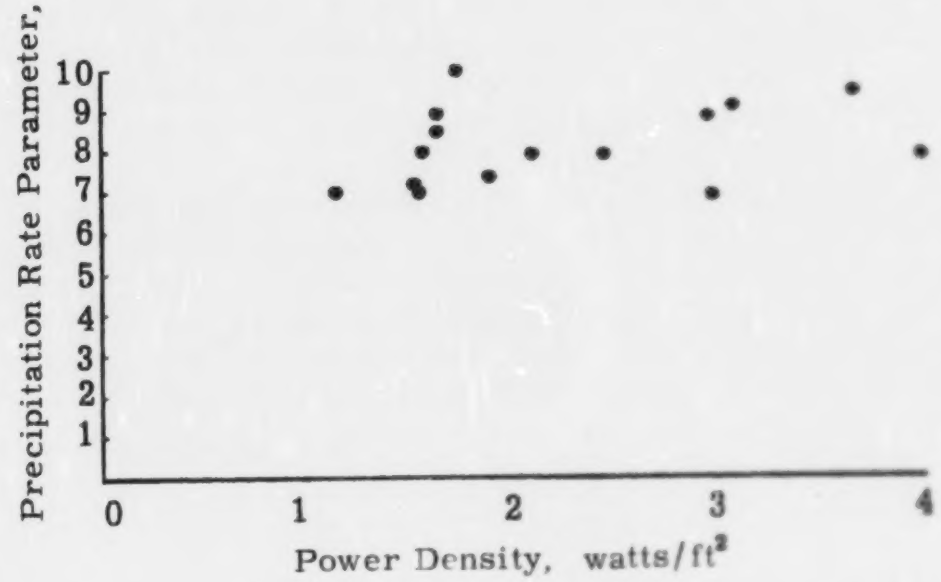

Figure 16.30. Relationship between Precipitation Rate Parameter and Power Density.

The significance of the rather limited range of precipitation rate parameters for pulp and paper is in the fact that the uncertainty in the performance of a new installation is significantly reduced over that for an electric utility installation. Figure 16.31 shows the variation in collection efficiency that could be expected for the range of uncertainty in the precipitation rate parameter reported from 11 installations. If one selected a precipitation rate parameter of 8 cm/sec and designed for a collection efficiency of 97%, the minimum collection efficiency attained for data from these 11 installations would be about 93%.

Having established the square feet of collecting electrode area required, the designer proceeds to select a specific precipitator size consistent with other practical aspects, such as: space limitations, if any; flue openings consistent with good flue configurations; practical ratios of plate length to height; and number of fields and sections required.

Most manufacturers attempt to specify a so-called standard precipitator, i.e., one consisting of standard collecting plate sizes, duct sizes, etc.

Electrical equipment. Equipment manufacturers through past experience develop confidential power requirements in terms of power input to the precipitator for the efficiency required. Most generally, these curves take the form of efficiency versus watts per 1000 cfm of gas to be treated.

The data shown in Figure 16.30 have a greater spread than for fly ash precipitators. The spread is due principally to the lack of measured input power data; some of the data shown are estimated power requirements supplied by the precipitator manufacturer, as opposed to field measured data. Minimum input power levels used as a rule of thumb are about 0.2 kW/1000 cfm for 90% efficiency, and around 0.8 kW/1000 cfm for 99.9% efficiency. Installed power supply capacity is generally larger than the estimated power input to accommodate changes in voltage-current requirements due to variation in dust and gas properties.

The degree of sectionalization also enters into specifying electrical equipment. In general, the degree of sectionalization is kept at a minimum depending on the difficulty of the problem and economic considerations.

Electrical equipment is, therefore, specified as a balance among the following: the input power required for a given efficiency; the degree of sectionalization required for the specific application; and economic considerations.

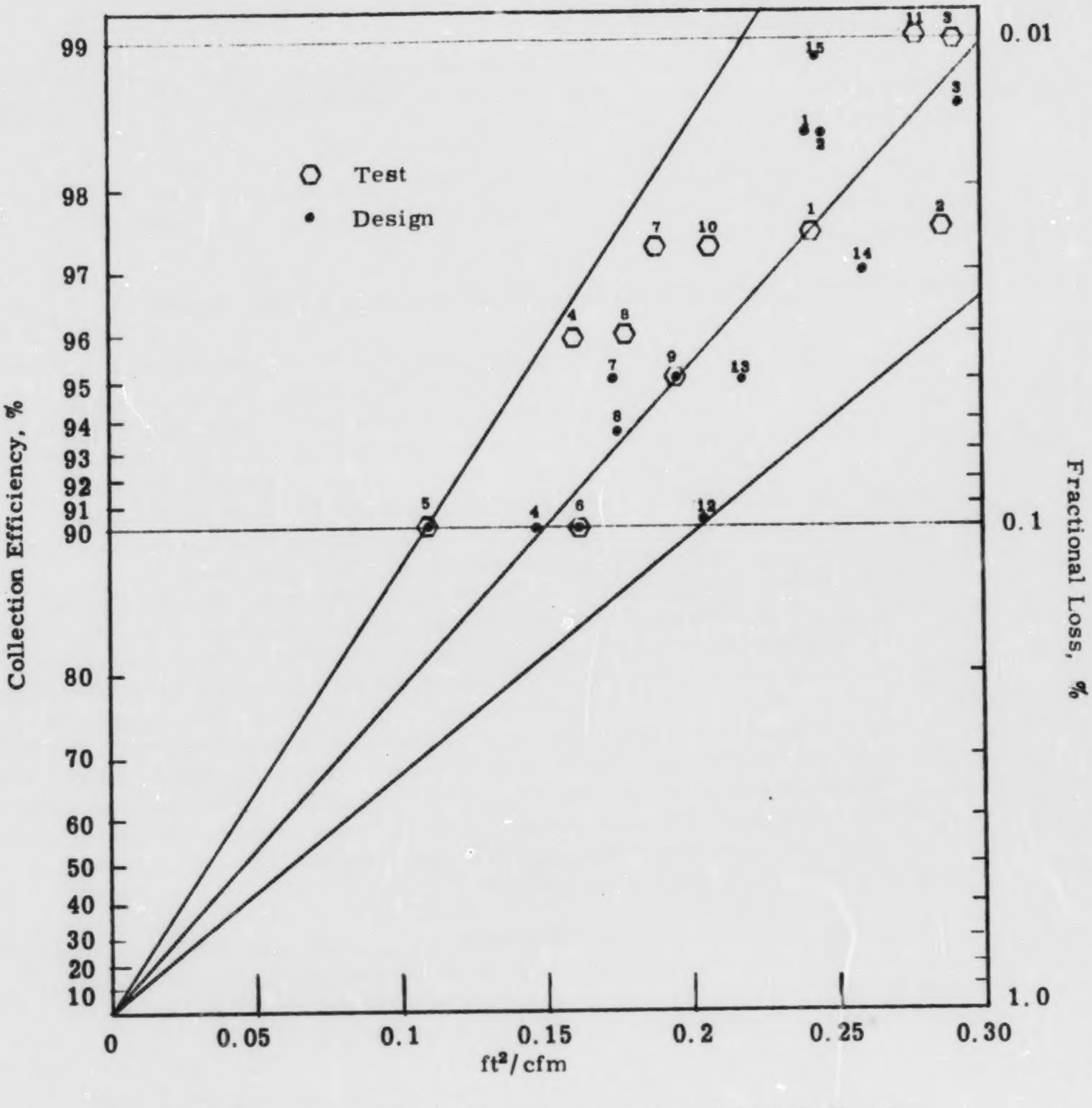

Figure 16.31. Relationship Between Design and Test Collection Efficiencies Vs. A/V Ratio for 11 Electrostatic Precipitators for Pulp and Paper Installations.

Electrode rappers. The type of collecting electrode rapper depends on the nature of the dust collected. In most instances, a heavy duty vibrator type rapper operating sequentially with the other rappers is specified for the salt cake application. The rappers are adjustable in vibration intensity and frequency of operation, and are generally field adjusted for maximum performance.

Dust removal system. Nearly all modern salt cake precipitators are equipped with the "wet bottom" type removal system. A few specify the dry drag bottom type.

16.8 GENERAL OBSERVATIONS AND DISCUSSION OF TRENDS

State of the art of the pulp and paper industry.[6,7,8,9] The present state of the art of the pulp and paper industry can be summarized as follows. Chemical recovery units now in operation handle black liquor from approximately 1500 tons of air-dried pulp per day, producing steam at 1500 psig and 925°F. They are usually the only chemical recovery units installed in the mills, and therefore must be capable of continued operation between scheduled mill shutdowns. In the period from 1965-1969, the average size recovery unit sold was for 545 tons per day of air-dried pulp, compared to about 150 tons per day in the 1939-1944 era. Recent emphasis on air pollution has resulted in requirements for elimination or drastic reduction of malodorous emissions, and 99+% removal of particulate from recovery unit flue gas. New "dry" type furnace systems have been developed which promise a high degree of odor removal by revised firing techniques, use of high efficiency economizers, and elimination of direct-contact of the hot flue gas with the black liquor fuel in the heat recovery steps. These systems may require precipitator changes such as dry-bottom dust removal equipment and higher efficiency to maintain the same level outlet particulate emissions as were achieved previously in the combination direct-contact evaporator and wet-bottom precipitator.

One of the problems associated with the use of electrostatic precipitators on paper mill recovery boilers is the occasional emission of light sulfate particles or flakes. The condition is called snowing. The particles settle out rather rapidly in the vicinity of the plant.

The condition of snowing can be caused by sudden release of particles accumulated in the duct work, by heavy rapping, by high gas velocities, or by overload of the precipitator by increased recovery boiler throughput.

Use of low energy scrubbers following the electrostatic precipitator is sometimes used to eliminate the snowing problem.

Precipitator trends. Trends in the use of electrostatic precipitators for the control of particulate emissions from the pulp and paper industry show a continuing increase since about 1945. This increase, shown graphically in Figure 16.32 is caused by two factors; the increase in pulp and paper production and the increase in the requirement for particulate control brought about by air pollution control legislation. This trend is expected to continue for the next several years.

Trends show an increase in the ability of manufacturers to predict the actual performance of precipitators for pulp and paper installations. A measure of the ability to predict this performance is by the ratio performance to design of the precipitation rate parameter. Table 16.6 and Figure 16.33 show an increase in the percentage of installations with an R value of 1 in the more recent years.

The significance of an R value of 1 is that the precipitator is neither too large nor too small. Thus, the purchaser should be able to effectively control the emissions without the need to purchase more precipitator than actually required. An R value greater than 1 means that the precipitator size is larger than actually required, while an R value less than 1 means that the performance guarantee was not met.

The design efficiency for the time that economic and operating data are available increased from an average of about 96.2% in the 1960-64 period to an average of about 98.4% in the 1965-69 period; current specifications often demand efficiencies in excess of 99%. When viewed in terms of particulate emission, assuming a constant inlet loading, the loss has been reduced on the average by about 87%. If all other factors remain constant, theory predicts an increase in the size of the precipitator by a factor of about 1.6.

Current trends in the design of the recovery section include the discontinuation of the practice of moisture reduction by direct contact between the black liquor and the hot flue gas. This direct contact process tends to increase the odor emissions from the installation. A second effect of the direct contact process is to control particulate emissions since the black liquor droplets provided a wet scrubbing action. With the removal of this mechanism of particulate emission control, increased collection efficiency will be required in the electrostatic collector.

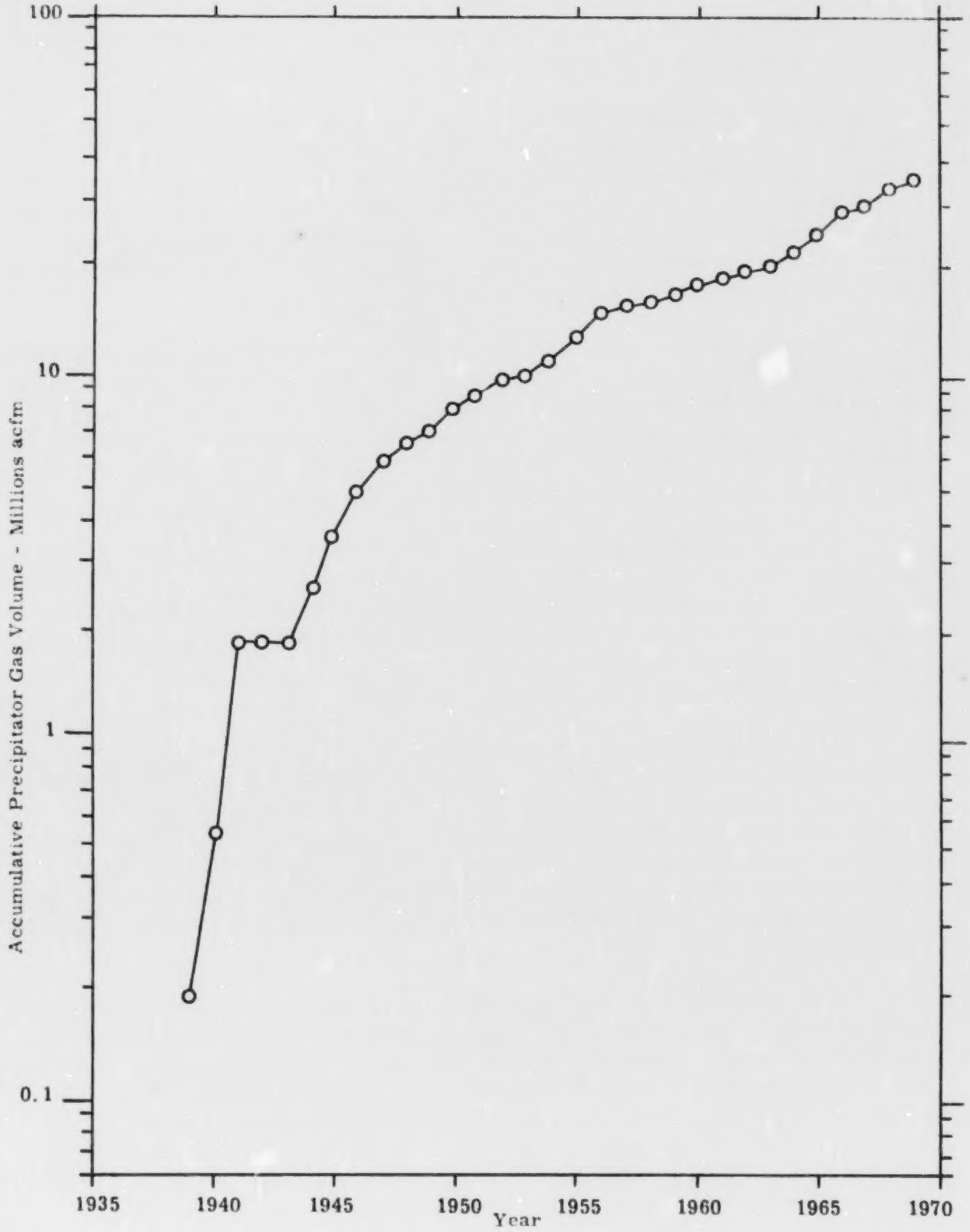

Figure 16.32. Accumulative Distribution of Precipitator Gas Volume (1939-1969).

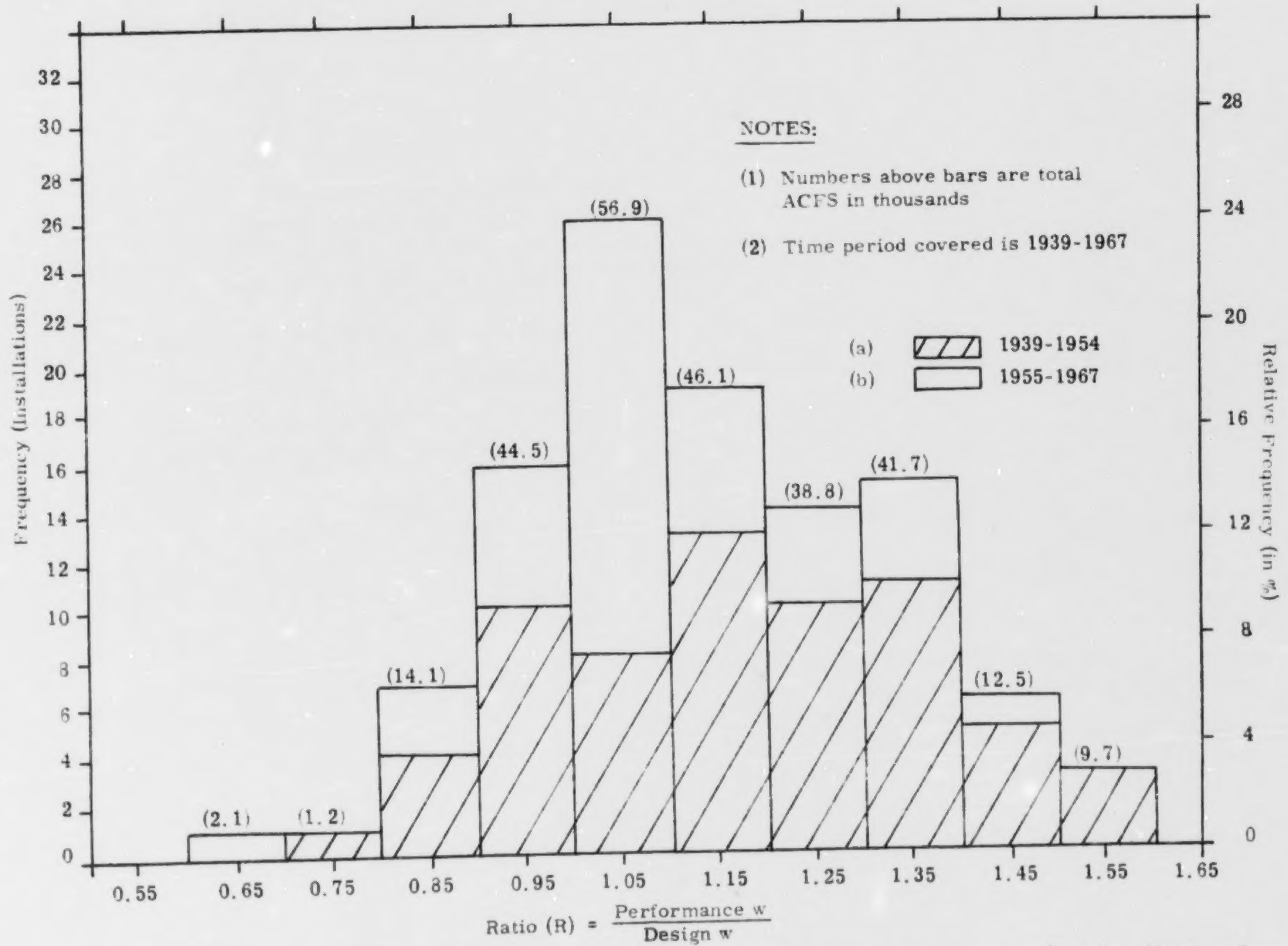

Figure **16.33**. Comparison of Actual Performance to Design Performance (Basis: Ratio of Migration Velocities Calculated Using **Deutsch Equation**).

Table 16.6

Precipitator Costs
(1965 through 1969)

Efficiency Range	Gas Volume Range - Millions acfm: < 0.100 FOB	< 0.100 Erected	0.100-0.299 FOB	0.100-0.299 Erected	0.300-0.499 FOB	0.300-0.499 Erected	> 0.500 FOB	> 0.500 Erected
99.0+	No Data	No Data	**117** (5)	178 (21)	No Data	**121** (5)	No Data	No Data
97.0 - 98.9	No Data	189 (1)	**70** (1)	151 (7)	No Data	**102** (2)	No Data	**94** (2)
90.0 - 96.9	No Data	No Data	No Data	91 (1)	No Data	**91** (2)	No Data	No Data

Note: (a) Costs are cents/acfm

(b) Numbers in parenthesis are sample size i.e., number of installation contract prices averaged to obtain indicated costs/acfm (all precipitator manufacturers represented).

The tendency to install larger recovery boilers also tends to increase the demand for electrostatic precipitators. In the past, high energy venturi scrubbers were used, in some instances, for controlling particulate emissions in the smaller units. The large units are almost exclusively controlled by electrostatic precipitators.

There is also a trend toward the increased production of bleached pulp. Bleached pulp operations require a larger precipitator than do the unbleached ones.

The above factors point to an increased use of larger electrostatic precipitators in the pulp and paper industry through the foreseeable future.

CHAPTER 16
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CHAPTER 17
ELECTROSTATIC PRECIPITATORS IN THE IRON AND STEEL INDUSTRY

17.1 INTRODUCTION

The applications of electrostatic precipitators in the iron and steel industry have been principally in the processes of iron production, coking, sintering, steelmaking, and iron melting in foundry cupolas.

Iron was first produced in this country during the Colonial Period (1645-1700) by the reduction of iron ore in charcoal furnaces. In 1834, a process utilizing what is termed a hot blast furnace was introduced. This was the forerunner of the modern blast furnace, which is still the primary process used in the reduction of iron ore.

The invention of the Bessemer converter in England in 1856 was a milestone in the production of steel. Following its introduction in this country, production of steel by the Bessemer converter steadily increased until in 1867, when production reached 3000 tons. The openhearth process, invented in England in 1868, soon began to replace the Bessemer converter until in the 1950's, when over 90% of the steel produced in this country was by the openhearth process. The electric furnace, introduced in the United States about 1905, provided a third means of steel production, and this method has continued to increase in importance.

Early in the 1950's, the top blown converter, basic oxygen furnace (BOF), or LD process, was introduced, and because of its high production rate, has rapidly become a major steelmaking process. Estimates for 1980 show about 25-30% of the steelmaking will be carried out in electric furnaces, 65-70% by the basic oxygen process, and 5% by the open hearth process. The Bessemer converter has essentially disappeared as a method of steelmaking.

Figure 17.1 is a flow chart showing the steps in the steelmaking process from the basic raw material and scrap input to the finished product. In a typical steel plant, there are about 10 major areas in which dust or fume is generated, and for which dust control equipment is required. A discussion of each of these areas is given in the subsequent sections of this chapter.

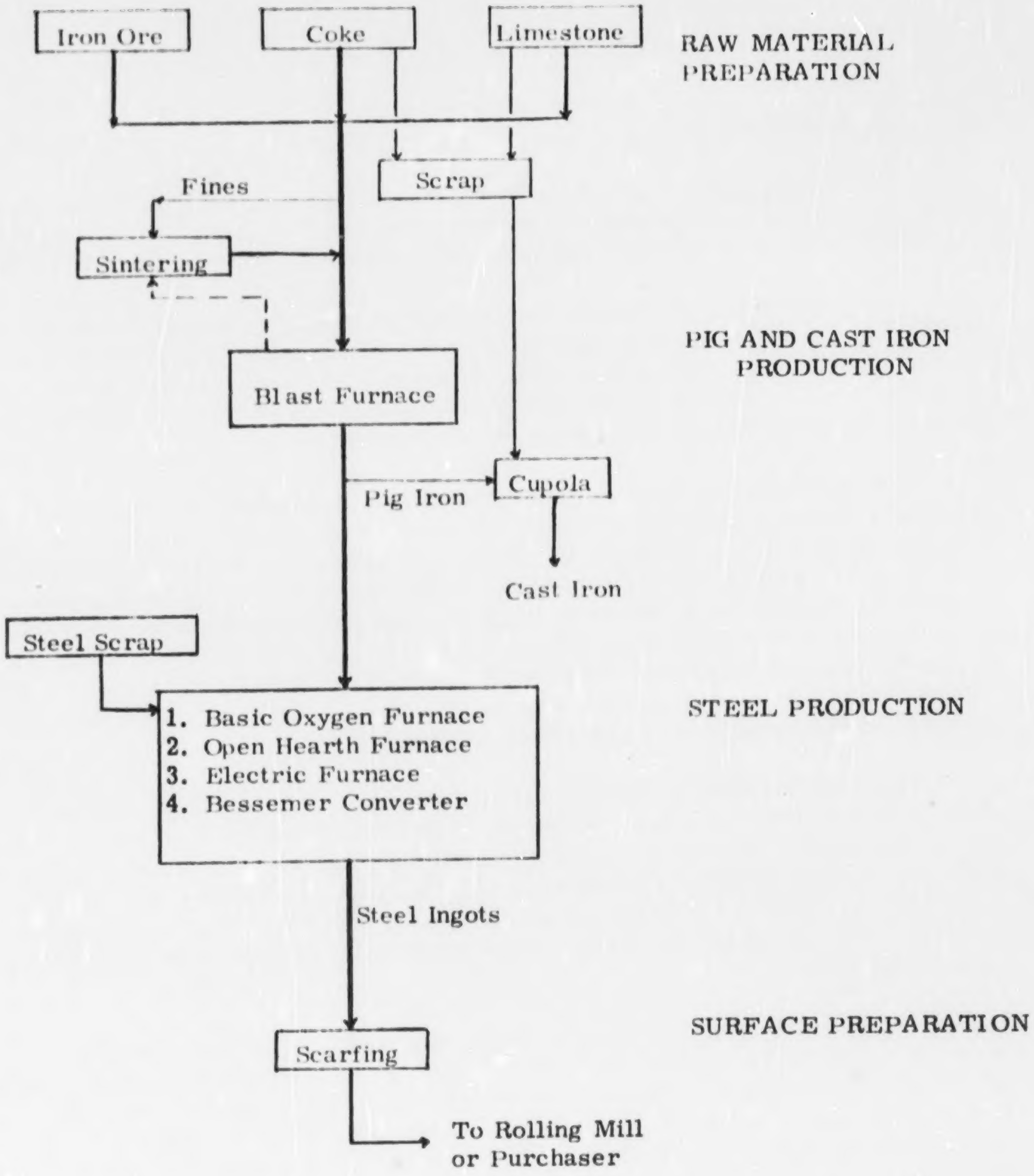
Iron Ore
Coke
Limestone
RAW MATERIAL PREPARATION
Scrap
Fines
Sintering
PIG AND CAST IRON PRODUCTION
Blast Furnace
Cupola
Pig Iron
Cast Iron
Steel Scrap
1. Basic Oxygen Furnace
2. Open Hearth Furnace
3. Electric Furnace
4. Bessemer Converter
STEEL PRODUCTION
Steel Ingots
SURFACE PREPARATION
Scarfing
To Rolling Mill or Purchaser

Figure 17.1. A Flow Chart Showing the Steps in the Steelmaking Process From the Basic Raw Material and Scrap Input to the Finished Product.

Figure 17.2 shows the rates of steel production by the various processes, and the total steel production for the period 1920 to 1969, with projection for the period 1969-1980. Figure 17.3 shows the production of steel, pig iron, and steel scrap used during the period 1920-1969.

Raw materials handling and preparation. The raw materials used in the production of iron are iron ore, limestone, and coal. These products are normally shipped by rail or barge to the blast furnace. The problems associated with the crushing and handling of iron ore and limestone are typical of those in quarrying and handling of bulk materials. Electrostatic precipitators are not normally associated with control of dust from these operations.

Before being charged into the blast furnace, the coal must be coked. This operation can produce particulate, as well as gaseous, emissions. Coking is a satellite operation that can be carried out on the premises of the steel plant or at remote locations.

The use of ore beneficiation to improve the iron content of the feed and the utilization of other fines has led to processes for pellitizing, nodulizing, and sintering of iron ore dusts and other fines before charging into the blast furnace.

17.2 APPLICATION TO COKE OVENS

The two methods of producing metallurgical coke are by the beehive and by-product processes. Since about 1918, the use of by-product processes has increased primarily because of the favorable economic aspects. The by-product process is presently used in the production of about 98% of all coke.

In the beehive process,[1,2] the coke oven is a dome-shaped structure, usually 10-15 ft in diameter, and 6-12 ft tall. It has a flat bottom which slopes toward an opening where the coke is discharged and where the air input is regulated during the coking cycle. The coal is charged into the oven through an opening in the roof, which also serves as a vent for the products of combustion and destructive distillation. The charge usually consists of about 6 1/2 tons of coal and the coking time averages 6-7 hours.

During the coking cycle, the pollutants from the beehive oven include smoke, dust, ash, hydrogen sulfide, phenols, and a wide variety of

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[1] Refer to the bibliography for this chapter.

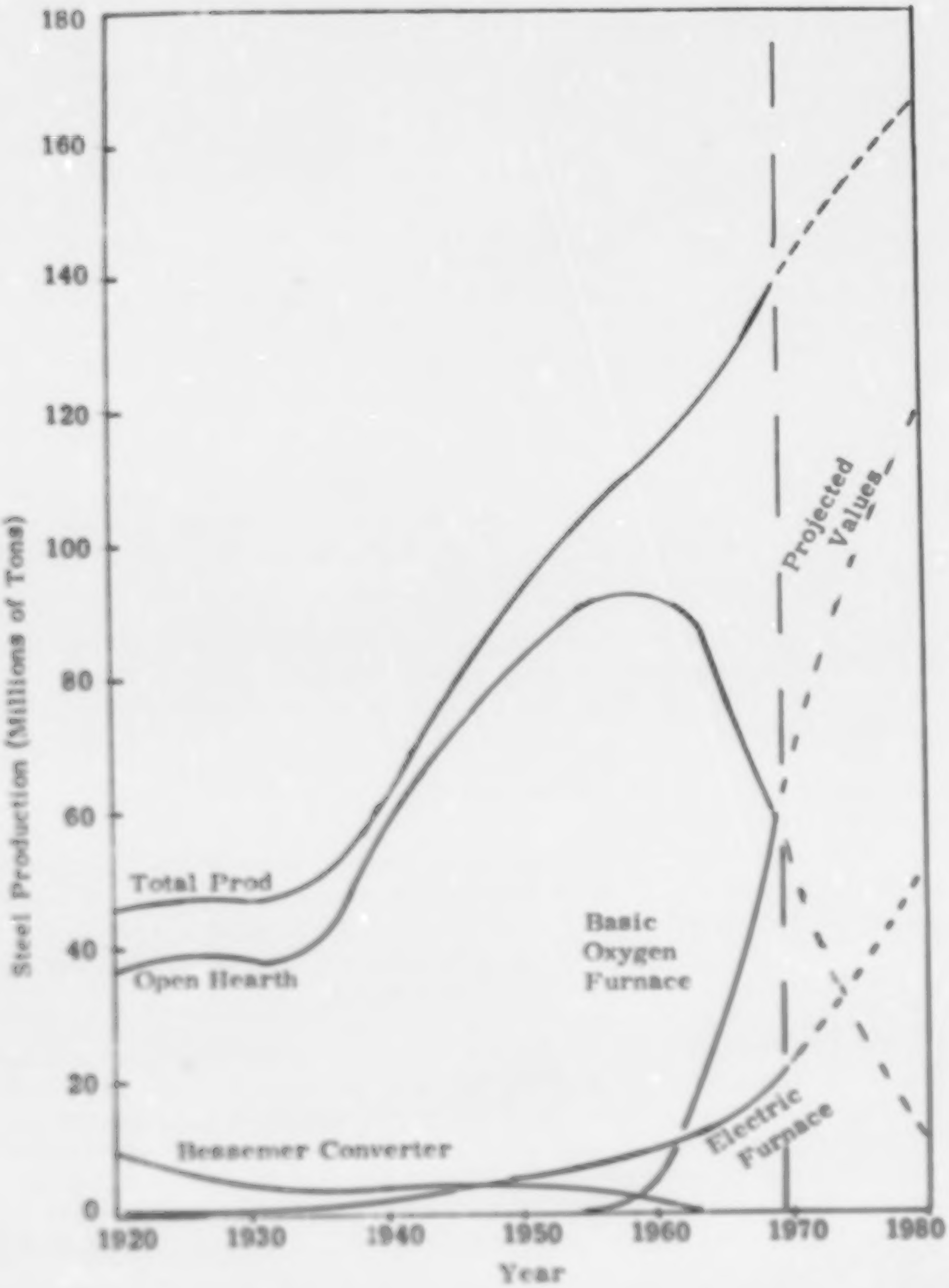


Figure 17.2. Rates of U.S. Steel Production by the Various Processes and the Total Steel Production for the Period 1920 to 1969, With Projection for the Period 1969-1980.

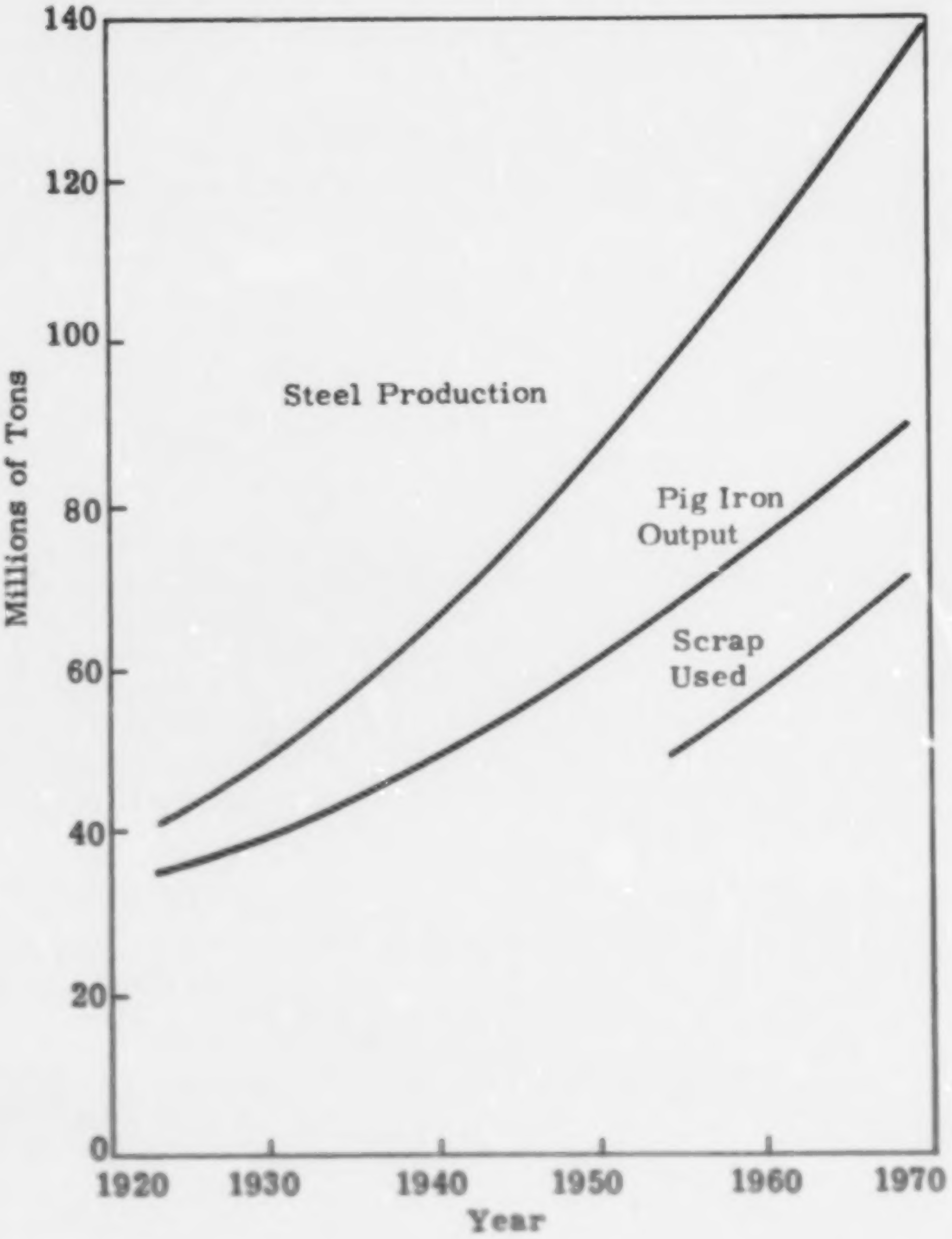


Figure 17.3. The Production of Steel, Pig Iron, and Steel Scrap Used During the Period 1920-1969.

saturated and unsaturated hydrocarbon compounds. The heavier pollutants, including particulate and heavy hydrocarbons, tend to fall out in the immediate area of the ovens, while the lighter hydrocarbons and particulate matter may be widely distributed.[3,4] Smaller amounts of particulate material are developed during quenching of the coke, but these are generally insignificant compared to that evolved during the coking period.

By-product coke ovens are usually 30-50 ft long, 6-14 ft high, and 12-22 in. wide. These are refractory brick structures and may be aligned in banks of 10-100 ovens per bank.[2] The ovens are charged through openings in the roof, and are discharged by a pusher rod operating through the length of the furnace. Both ends of the ovens are equipped with self-sealing doors, which help to restrict emissions from these locations during coking. Heating chambers are lined alternately with the ovens and are fired with preheated air and a fuel, such as natural gas, coke oven gas, or blast furnace gas. Charges average 16-20 tons of coal per oven, and the coking time is usually 16-20 hours.[5]

General aspects of particulate and gaseous emissions. In the beehive coking process, all of the products of combustion and destructive distillation of the coal are vented directly to the atmosphere. These emissions consist of smoke, dust, and a wide variety of organic materials. An average of 25% of the weight of the coal charged into the coke ovens is emitted as gaseous and particulate material. From 10,000-12,500 std cu ft of gaseous emissions are developed per ton of coal in the coke oven. The actual volume of gaseous emissions depends, to some extent upon the grade of the coal, temperature of the oven, and the coking time.

The gases and particulate emissions from by-product coke oven retorts are essentially the same as those from beehive ovens, however, the products of destructive distillation in this process are withdrawn from the oven and processed to produce useful by-products.

The carbonization of 2000 lbs of coal results in the following products:

coke	1200 - 1400 lbs
coke breeze	100 - 200 lbs
coke oven gas (550 BTU/ft^3)	9500 - 11500 cu ft
crude light oil	2 - 4 gal
tar	8 - 12 gal
ammonium sulfate	20 - 28 lbs
ammonium liquor	15 - 35 gal

Atmospheric emissions may be produced during by-product coke operation during charging, discharging, quenching, and to a small extent--during coking, because of leaks around the oven doors. When the retorts are charged, fines may be emitted from the roof, and some volatile fractions evolved when coal is put into a hot oven.[6]

Removal of the particulate and tar from by-product oven gas was by means of "P and A" tar extractors and "shaving" scrubbers prior to the introduction of electrostatic precipitators. The "P and A" tar extractor was an impingement type scrubber which might achieve an efficiency of 60-70% with a pressure drop of 6-7 in. water. The "shaving" scrubber consisted of a filter of wood shavings, which provided an efficiency of 80-85% with a pressure drop of 10-12 in. water.

Electrostatic precipitators provide efficiencies of 95-99% with a pressure drop of around 0.5 in. water. Introduction of the electrostatic precipitator for removing tar from coke oven gas was delayed because of fear of igniting a combustible mixture from the electric spark inherent in electrostatic precipitator operation. The problem was handled by careful control of air inleakage to maintain the gas at noncombustible conditions.

The purpose of gas cleaning in a by-product coke oven is to remove the particulate and heavy tar from the gas prior to recovery of various hydrocarbons. The equipment for accomplishing this is shown in Figure 17.4. The hot gas leaves the ovens at 1100 to 1300°F where it is cooled to about 95°F. This cooling process condenses out a considerable amount of the tar and ammonia liquor. These condensed products are processed further in other equipment for separation and refinement.

The gas then passes through the exhauster and enters the electrostatic precipitator or detarrer. Here the suspended oil and tar are precipitated from the gas stream. Efficiency of collection is normally between 95 and 99% depending upon requirements.

Gas then moves through reheaters, ammonia absorbers or saturators, gas coolers, and light oil scrubbers, before entering the gas holder. If the gas is to be used for underfiring the coke ovens, the gas must be further cleaned in another electrostatic precipitator referred to as a fuel gas precipitator. This precipitator cleans the gas sufficiently to prevent the deposition of tar in the very small apertures of the coke oven burners. Some of this finely cleaned gas is also used to ventilate the insulator compartments of the primary precipitator. The gas volume used for underfiring is usually a small fraction of the total gas. Consequently, only

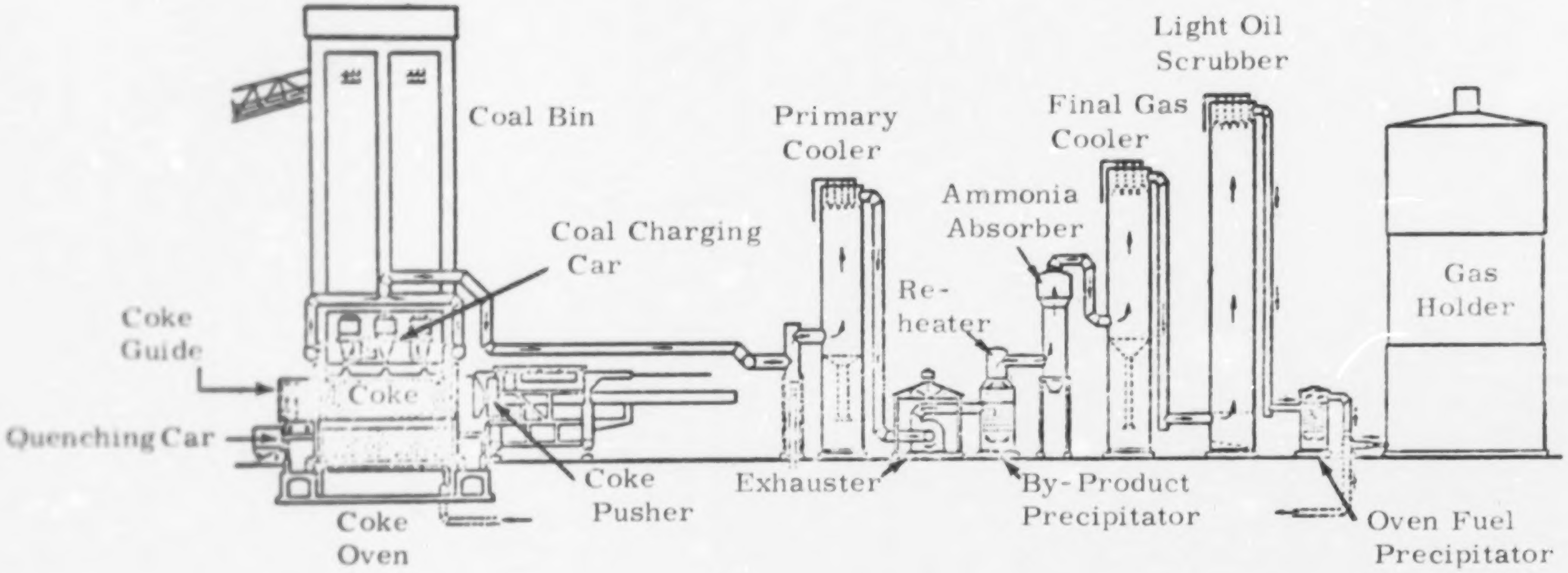

Coal Bin
Coal Charging Car
Coke Guide
Coke
Quenching Car
Coke Oven
Coke Pusher
Primary Cooler
Exhauster
Re-heater
Ammonia Absorber
By-Product Precipitator
Final Gas Cooler
Light Oil Scrubber
Gas Holder
Oven Fuel Precipitator

Figure 17.4. Equipment Used to Remove the Particulate and Heavy Tar from Coke Oven Gas Prior to Recovery of Various Hydrocarbons.

relatively small precipitators are required.

The analysis of the gas coming from the coke oven is approximately as follows:

Constituent	Volume, %
CO_2	2.2
O_2	0.8
N_2	8.1
CO	6.3
H_2	46.5
CH_4	32.1
C_2H_4	3.5
C_6H_6	0.5

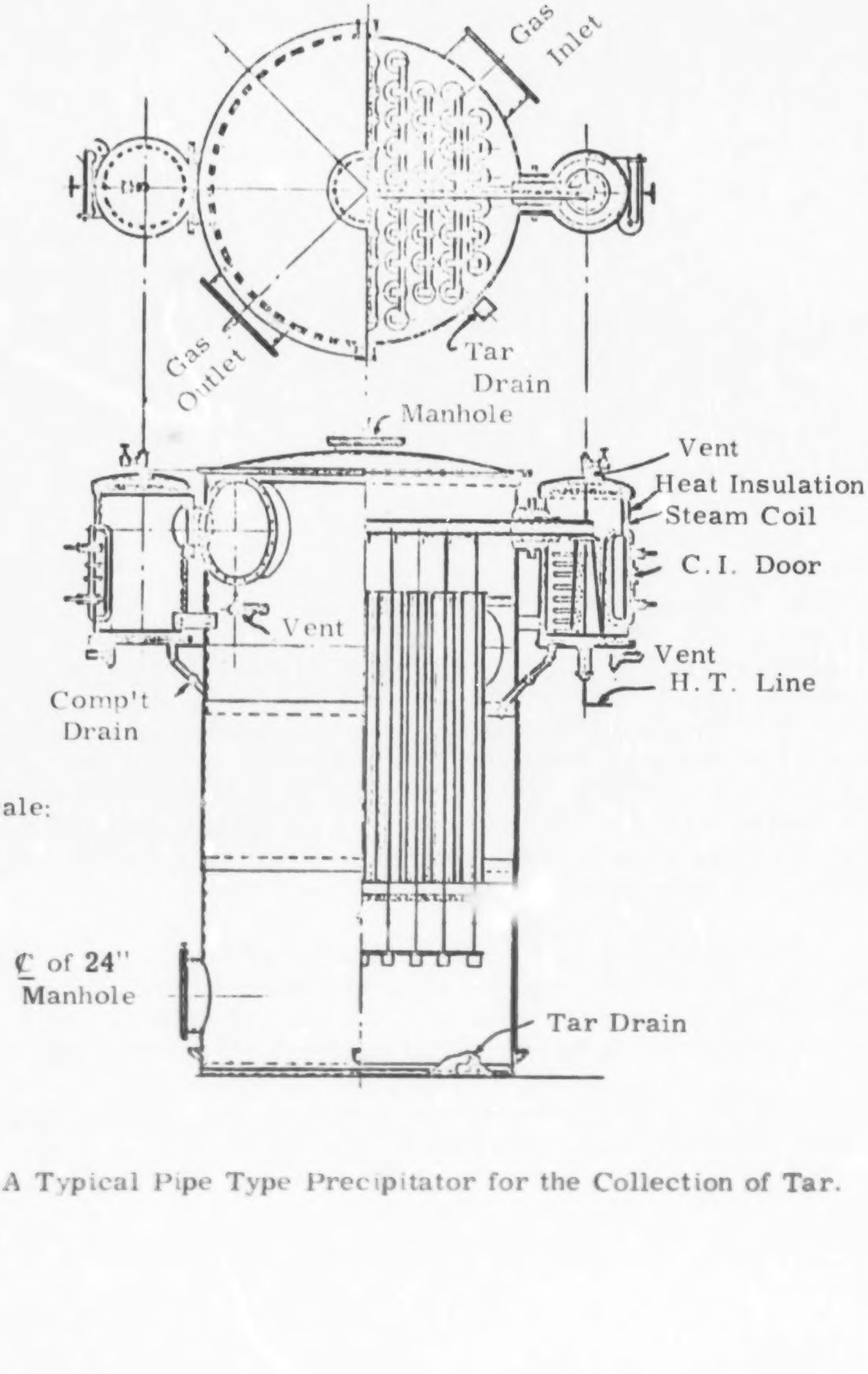
The heat value is more than 500 Btu/cu ft.

The concentration of suspended matter at the precipitator inlet varies from 1 to 15 grains/scf. Water comprises approximately 50% of total precipitate weight, the remainder consists of tar, oil, etc.

Description of electrostatic precipitator for detarring. The precipitator most often used for detarring consists of a group of grounded pipes 6 in. to 8 in. in diameter and 6 ft to 9 ft in length, suspended from a header plate in a round shell. A discharge electrode wire is suspended axially in each pipe. A potential difference of 35-50,000 volts is established between the discharge and collecting electrodes. The dirty gas enters the lower header and is passed through the pipes. Figure 17.5 shows a typical design for this type of precipitator.

Figure 17.6 shows a self-contained tar precipitator. This design utilizes a 35 kV electrical set with tube rectifier and controls housed within a steel compartment attached to the precipitator shell. The collecting pipes of this unit are 6 in. in diameter and 6 ft long.

Another precipitator design used for cleaning coke oven gas is shown in Figure 17.7. The collecting electrodes in this design are concentric tubes. The discharge electrode wires are centrally located in the annulus between the tubes. Both surfaces of the tubes serve as collecting surface. The electrical equipment is similar to that used with pipe units.


Gas Inlet
Gas Outlet
Tar Drain
Manhole
Vent
Heat Insulation
Steam Coil
C.I. Door
Vent
H.T. Line
Vent
Comp't Drain
℄ of 24" Manhole
Tar Drain

Approximate Scale:

5 ft = 1 in.

Figure 17.5. A Typical Pipe Type Precipitator for the Collection of Tar.

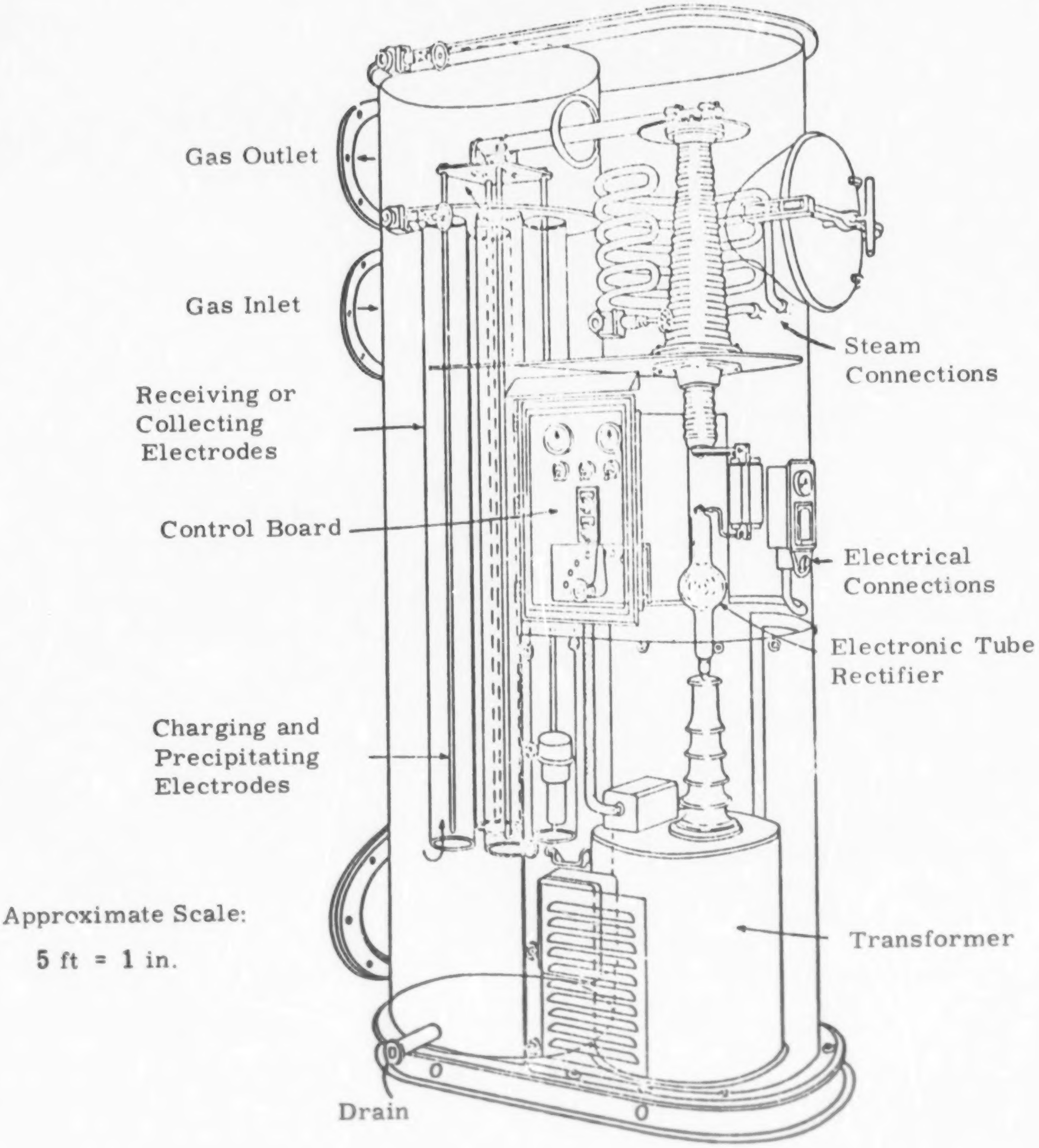

Gas Outlet
Gas Inlet
Receiving or Collecting Electrodes
Control Board
Charging and Precipitating Electrodes
Approximate Scale:
5 ft = 1 in.
Drain
Steam Connections
Electrical Connections
Electronic Tube Rectifier
Transformer

Figure 17.6. An Integral Tar Pipe-Type Precipitator.

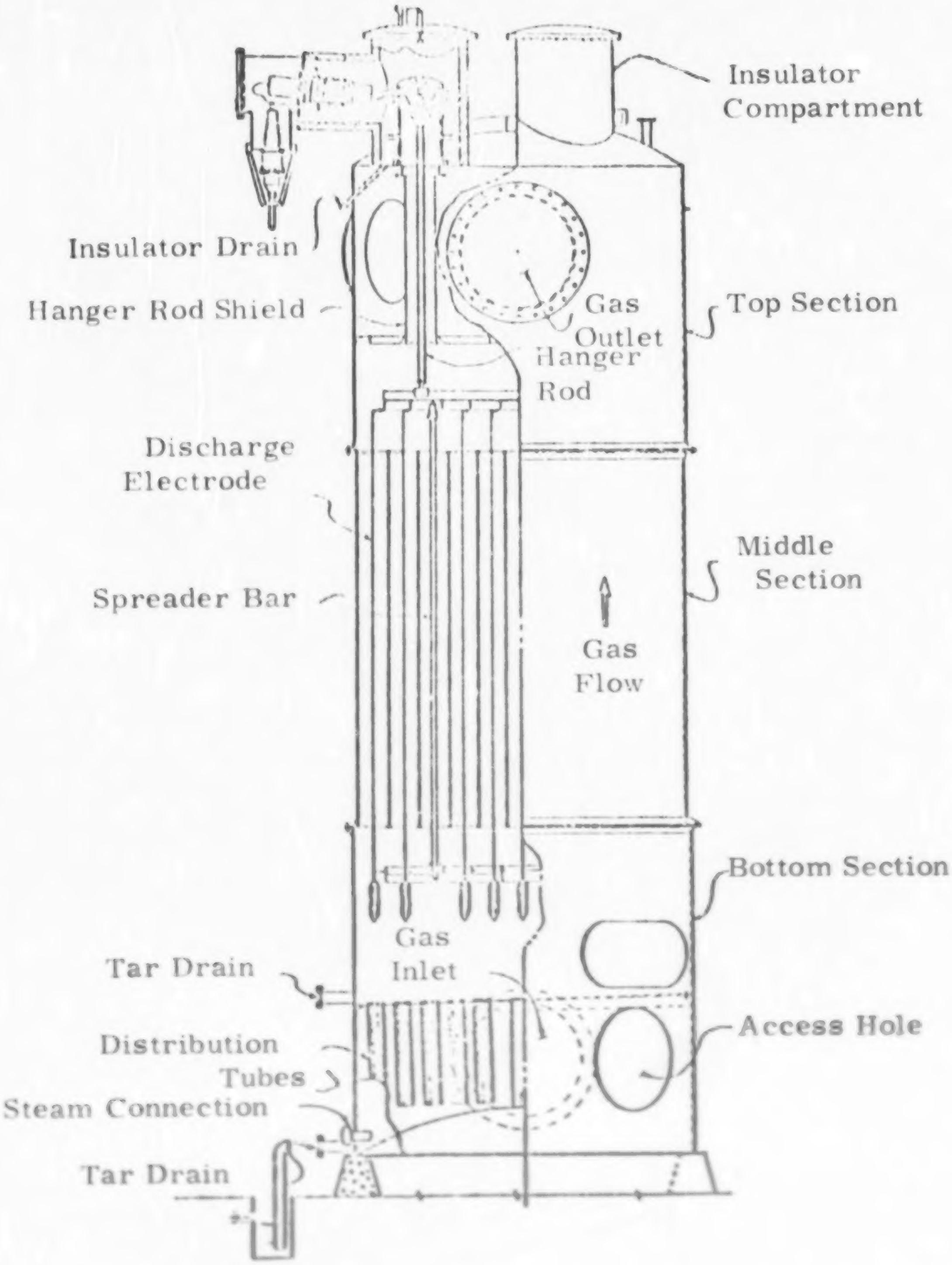

Insulator Compartment
Insulator Drain
Hanger Rod Shield
Gas Outlet
Top Section
Hanger Rod
Discharge Electrode
Middle Section
Spreader Bar
Gas Flow
Bottom Section
Gas Inlet
Tar Drain
Access Hole
Distribution Tubes
Steam Connection
Tar Drain

Figure 17.7. Concentric Ring Detarrer.

Operational problems and their solution. The use of the electrostatic precipitator for the cleaning of coke oven gas has been successfully applied since the solution of the three basic problems inherent in the application. These were electrical insulation, corrosion and safety.

The saturated gas and fluid tars make the electrical insulation particularly difficult, and require the following corrective measures. First, the insulators are located out of the gas stream. Second, the insulators are kept heated with steam coils and thermal insulation is applied to the insulator compartments. Third, regular maintenance programs must be followed to clean the insulators at programmed intervals. Water seals have been installed on many units to permit access to the insulators for replacement or cleaning. This reduces the time and expense required for insulator maintenenace. The fourth measure used is the ventilation of the insulator compartments with clean gas to prevent dirty gas from entering the compartments.

Corrosion has not generally been a serious problem in this precipitator application, although high sulfur coking coals have caused severe corrosion in some instances. Low tar content in the gas to the precipitator has also created corrosion problems.

Corrosion, when it occurs, is most severe on the collecting electrodes and in the insulator compartments. The corrosion of the collecting pipes occurs on the outside of the pipes.

The inside surfaces appear to be protected by the tar film which is created by the precipitation of the oil and tar suspended in the gas. Protection to the outside surfaces of the electrodes is provided on many installations by periodically spraying the areas with warm tar. The internal piping and nozzles for this operation are provided in the original design. The corrosion of the insulator compartments is combatted in two ways: the inside surfaces of the compartments are often covered with a protective coating, and the use of clean gas flowing into the compartments prevents the entrance of corrosive gases.

The gases are combustible and there is occasional sparking in the precipitator. With this condition, of course, any leakage of air or oxygen into the system could result in an explosive mixture and a destructive explosion. To eliminate this possibility, the design must prevent air inleakage and provide for adequate purging. Most, but not all, coke oven units are located after the exhauster (Figure 17.4) to assure operation under a positive pressure.

Design considerations. Based on data from a group of installations covering a period of 1931 to 1961, design efficiencies are generally about 95-99%. Data showing design parameters such as the surface area, gas volume, installed power, etc., are summarized in Table 17.1 Typical design parameters for coke oven gas precipitators are given in Table 17.2 as an illustrative example.

Cost data. Cost data on coke oven gas precipitators is generally poor because of the age of the installations. Data that are available show such a wide range ($1.10 to $6.00 FOB and $1.40 to $7.00 erected costs per cfm) as to be of little value.

Trends. A review of the installation of coke oven gas precipitators shows that this application has decreased substantially in the past 10 years.

17.3 APPLICATION TO SINTER PLANTS

There are two commercial processes for agglomerating iron-bearing fines to produce an improved blast furnace burden. These are sintering and pelletizing. The sintering process is used primarily for blast furnace flue dust, mill scale, and other metallurgical fines collected during steel making processes. Pelletizing is usually associated with the agglomeration of beneficiated, low grade iron ores.

The sintering process is a continuous operation performed on interconnected grates which form a slow moving loop. The beds formed by these grates are usually 8-12 ft wide and 90-100 ft long.[7] Raw materials, consisting of iron-bearing fines, coke or coal dust, and a fluxing material (limestone or dolomite), are thoroughly mixed and put on the grate. As the mixture moves through the stand, the bed is ignited by surface burners, and combustion is sustained by drawing air through the coal-bearing bed with large fans. Combustion of the coal generates sufficient heat to raise the temperature of the mass to 2400-2700°F. This high temperature causes agglomeration of the mixture to porous, coherent lumps.[8] The material is then collected and stored for subsequent use in blast furnaces. Modern sinter plants have capacities of 1000-6000 tons of sinter per day.[2]

Table 17.1

Typical Coke Oven Gas Precipitator Data

Gas Volume	5,000 - 20,000 cu ft/min
Design Efficiency	95 - 99%
Precipitation Rate Parameter	0.2 - 0.55 ft/sec
Type Discharge Electrode	1/4 in. sq twisted wire
Type Collecting Electrode	8 in. dia steel cylinder
Voltage	50 kV - 70 kV peak
Average Electric Field	11.5 kV/in.
Average Grain Loading	0.3 - 0.6 gr/acfd
Average Gas Velocity	8.0 ft/sec
Power/Gas Volume Flow	100 - 200 w/1,000 acfm

Table 17.2

Typical Coke Oven Gas Precipitator Design Parameter

Example of Coke Oven Gas Precipitator Performance
Precipitator: 148 pipes, 8 in. dia x 9 ft long

Performance Data

v_g = gas flow = 16,500 cfm at 100°F

Inlet concentration = 0.68 gr/cf
Outlet concentration = 0.0068 gr/cf

Efficiency = 99%

$$\frac{A}{v_g} = \frac{148 \times \pi \times 8/12 \times 9}{16,500/60} = 10.2 \text{ sec ft}$$

$$w = \frac{v_g}{A} \ln \frac{1}{Q} = \frac{1}{10.2} \ln 100 = \frac{4.6}{10.2} = 0.45 \text{ ft/sec}$$

$$v = \text{gas velocity} = \frac{v_g}{A_c} = \frac{16,500}{60 \times 148 \times \pi \times (\frac{1}{3})^2} = 5.3 \text{ ft/sec}$$

$$P_c = \text{corona power} = 3100 \text{ watts}$$

$$\frac{P_c}{v_g} = \frac{3100}{16.5} = 190 \text{ watts}/1000 \text{ cfm}$$

Figure 17.8 is a sketch of a typical sintering machine showing the assembly of pallets which form the belt which conveys the charge through the sinter machine. The ignition furnace is either gas or oil-fired, and its purpose is to bring the fuel in the charge to its kindling temperature, after which the combustion of the fuel supplies the heat. The flue gas is collected in a multiplicity of compartments called wind boxes located along the length of the machine, from which it is transported through ducts to the dust collecting equipment, usually consisting of a combination mechanical-electrostatic dry type precipitator.

Particulate emissions from sinter machines. In the sintering process, particulate material may be emitted during handling of the raw material, combustion of coal mixed with the ore, or during screening operations. The amount and composition of the particulate and gaseous emissions depends on several factors, including the type of ore used, the efficiency of mixing, and the distribution of the unfired sinter on the grate. Typical ranges of composition are shown in Table 17.3.

Under normal conditions, the dust load varies from 5-100 lb of dust per ton of sinter produced, with a mean of around 20 lb of dust per ton of sinter.[8-11] Gas volumes exhausted usually vary between 100,000 and 450,000 cu ft per min,[11,12] with dust loadings of 0.5-6.5 grains/scf.[2,8,11,13] From 80-90% of the total particulate material from the sintering operations are greater than 20 microns in size by mass.

The size of the dust coming off the sinter machines is rather coarse; however, after passing through mechanical collectors, the dust burden can be quite fine, as shown in the photomicrograph, Figure 17.9. Typical particle size analyses for sinter machine dust before and after a mechanical collector, are shown in Figure 17.10. Particulate loadings to the precipitator typically vary between 0.1 and 1.0 gr/scf, although loadings up to 1.5 gr/scf are occasionally encountered.

Gas temperature can fluctuate between 150 and 400°F, but 200 to 300°F may be taken as a reasonable range that can be expected. Moisture in the gas can vary between 5 and 15% by volume and sulfur oxide content from 25 to 500 ppm depending upon the sulfur content of the charge.

The electrical resistivity of sinter machine dust can vary radically depending upon the type and amount of material used in the burden make-up. Figure 17.11 shows laboratory measurements made on dust entering an

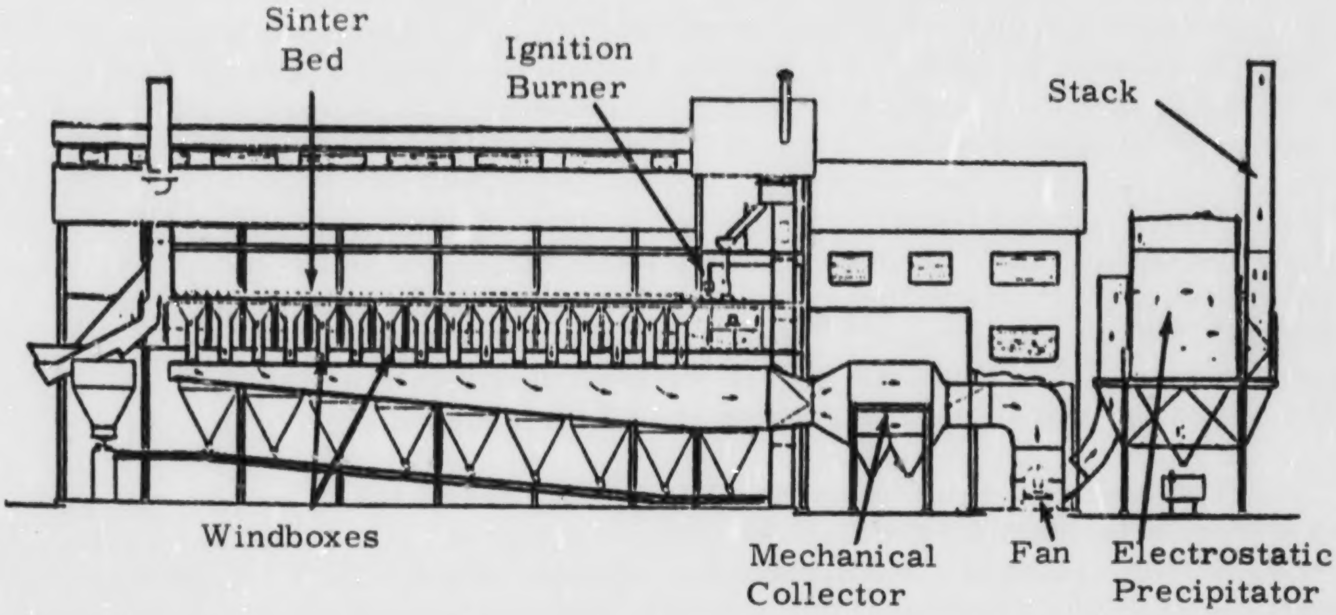

Sinter Bed
Ignition Burner
Stack
Windboxes
Mechanical Collector
Fan
Electrostatic Precipitator

Figure 17.8. Iron Ore Sintering Machine Gas Cleaning With Electrostatic Precipitator.

Table 17.3

Range of Composition of Particulate from Sinter Plants

	Weight %
Fe	Up to 50% and more
SiO_2	9 to 15
CaO	7 to 12
MgO	1 to 2
Al_2O_3	2 to 8
C	0.5 to 5
Alkalis	0 to 2
S	up to 2.5

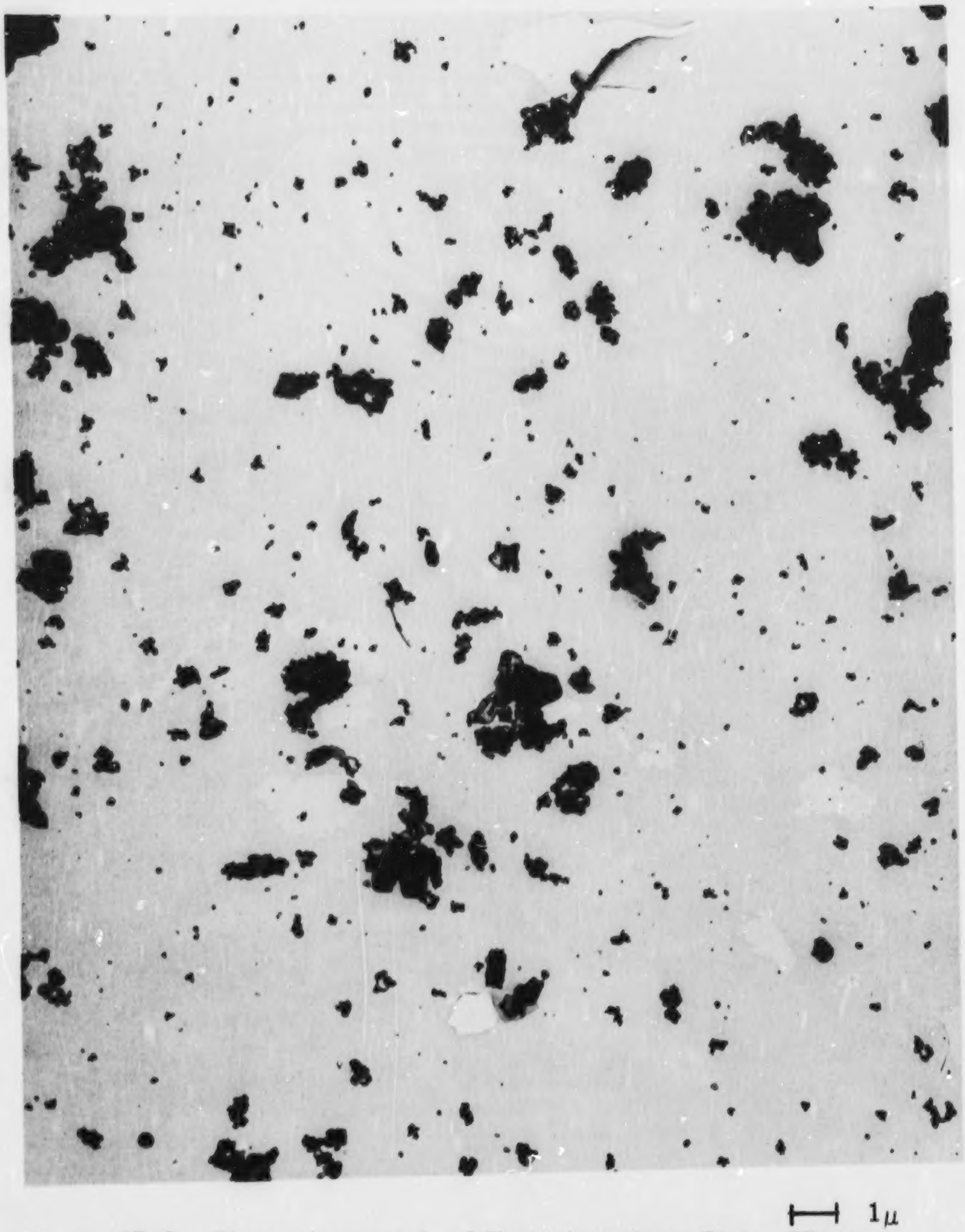


Figure 17.9. Photomicrograph of Emissions from Sinter Plants, 7800X, After Mechanical Collector.

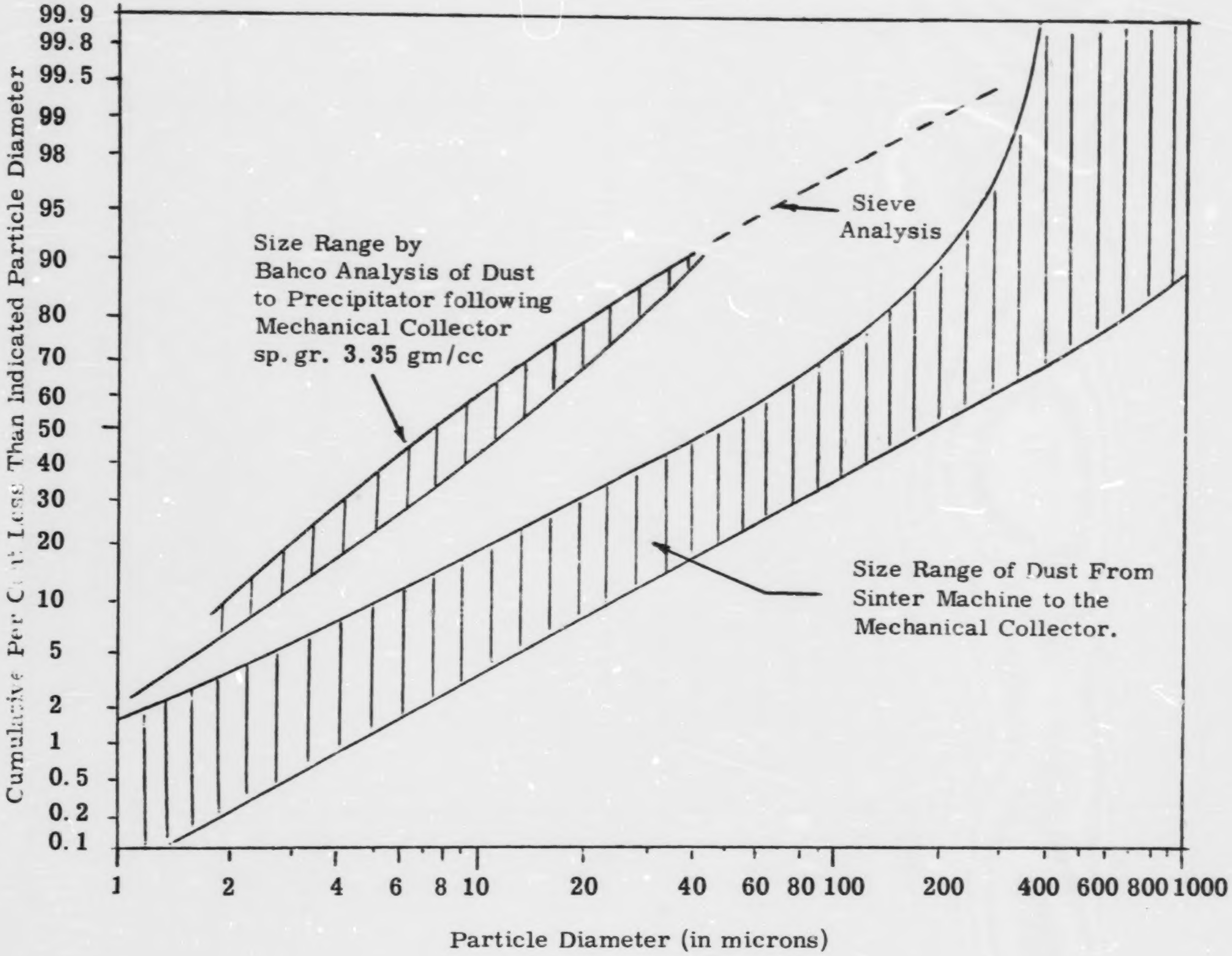
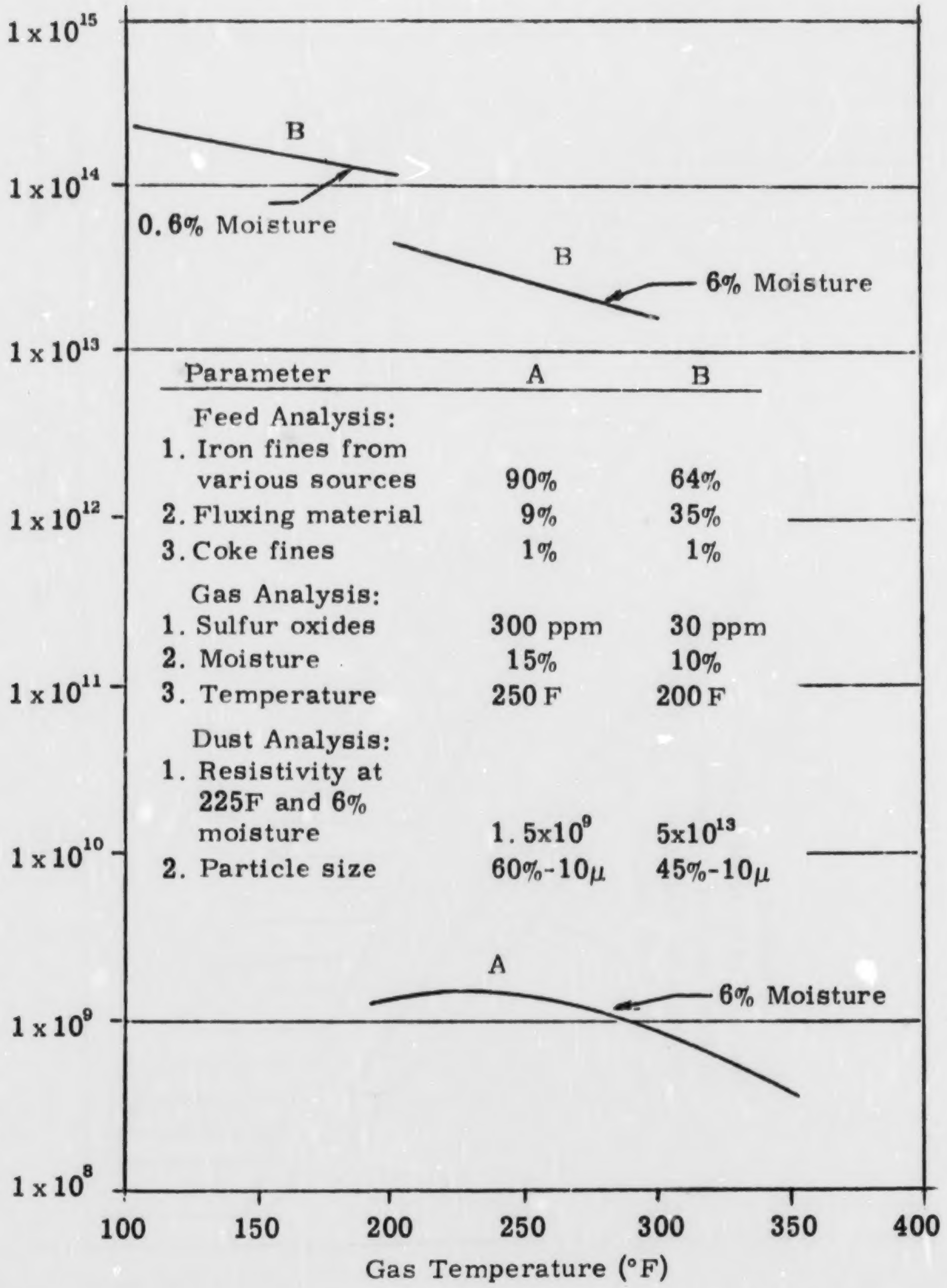


Figure 17.10. Particle Size Distribution By Weight of Sintering Machine Dust.



Parameter	A	B
Feed Analysis:		
1. Iron fines from various sources	90%	64%
2. Fluxing material	9%	35%
3. Coke fines	1%	1%
Gas Analysis:		
1. Sulfur oxides	300 ppm	30 ppm
2. Moisture	15%	10%
3. Temperature	250 F	200 F
Dust Analysis:		
1. Resistivity at 225F and 6% moisture	1.5×10^9	5×10^{13}
2. Particle size	60%-10μ	45%-10μ

Figure **17.11**. Electrical Resistivity of Sintering Machine Dust.

electrostatic precipitator from the same sintering machine with a lapse of about five years in time. The type burden charged in each case is indicated on the figure.

The temperature at which the gases enter the precipitator, as well as the moisture and sulfur content of the fuel, play an important role in establishing the resistivity of the dust. In the temperature range of 200-300°F, the volume conductivity is quite low. Consequently, conditions must be established for surface conduction if resistivity is to be maintained in the proper range for good precipitation.

The conditions for establishing a surface conductive film on the dust are not unlike those required for fly ash conditioning. Hence, moisture and sulfur content of the flue gas are important parameters.

Recent trends in the additions of limestone and dolomite to the sinter have altered the composition of the collected dust, and have increased the apparent resistivity, so that precipitation is difficult. The role of lime additions in altering resistivity has not been fully explored. Possible mechanisms are: (1) the lime combines with the moisture present to form a hydrate with high resistivity, and (2) the lime reacts with the sulfur present to form a high resistivity sulfate.

In either instance, the addition of large quantities of fluxing material alters the surface conduction properties, and hence, the dust resistivity.

Sinter machine precipitators. The precipitator used for sinter machine gas is typically a single-stage, horizontal-flow, duct-type unit. The shell and hoppers are fabricated steel, and may or may not be thermally insulated. Table 17.4 shows the number and installed capacity of sinter machine precipitators during the period 1951-1968. The design efficiency trends, also shown in the same table, indicate a higher collection efficiency for newer installations. Figure 17.12 shows the installed capacity of electrostatic precipitators for sinter machine dust collection for the 1951-1969 period.

Table 17.5 lists some important performance parameters for precipitators collecting sinter machine dust.

Table 17.4

Use Statistics on Sintering Machines, 1951-1968

Pptr. Contract Year	No. of* Install.	No. of Elec. Pptrs.	Total Gas Vol. (10^6 acfm)	Total Gas Vol. Accumul. (10^6 acfm)	Weighted Eff. on acfm Basis		Five-Year Period	
					Yearly	5-Yr.	Total Gas Volume (10^6 acfm)	Average Vol/Yr during Period (10^6 acfm)
1951	1 (2)	2	0.330	0.330	90.0			
1952	1 (3)	3	0.450	0.780	90.0			
1953	1 (2)	2	0.280	1.060	-	90.0	1.06	0.212
1955	1 (1)	1	0.440	1.500	95.0			
1956	2 (4)	4	1.330	2.830	90.0			
1957	1 (1)	1	0.457	3.287	98.0			
1958	2 (2)	2	0.735	4.012	-			
1959	2 (2)	2	0.592	4.604	92.1	92.5	3.554	0.711
1962	1 (2)	2	0.360	4.964	93.0			
1964	1 (1)	1	0.175	5.139	98.5	94.8	0.535	0.107
1966	1 (1)	1	0.194	5.333	96.5			
1967	3 (5)	5	0.982	6.315	97.1			
1968	2 (6)	3	1.190	7.505	97.0	97.0	2.366	0.473
Grand Totals	19 (32)	29	7.505					

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*Numbers in parenthesis are machines involved in installations indicated.

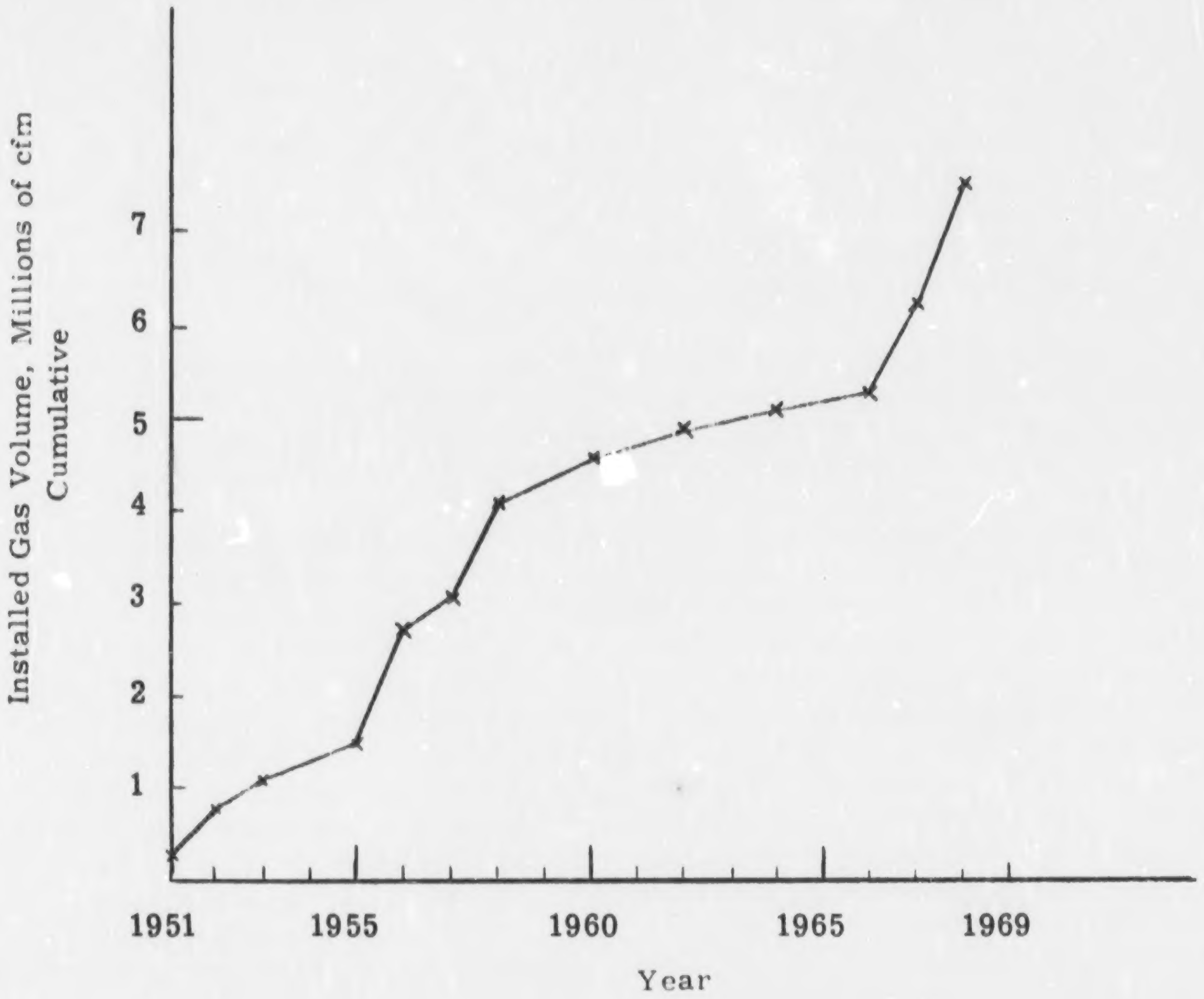


Figure 17.12. Installed Capacity of Electrostatic Precipitators for Sinter Machine Dust Collection for the Period 1951-1969.

Table 17.5

Performance Parameters for Precipitators Collecting Sinter Machine Dust

	Design	Test
No. of installations	5	2
Avg. gas velocity	4.3 ft/sec	
Avg. inlet gas temperature	245°F	245°F
Avg. electric field	8.1 kV/in.	8.1 kV/in.
Avg. dust loading gr/acf	1.0	0.75
Avg. precipitator power	71 watts/1000 cfm	

Design of electrostatic precipitators for sinter machine dust collection follows the general procedures outlined in Chapter 9, Part I. The size precipitator required is based on the collection efficiency, gas volume, and the precipitation rate parameter, as defined by the Deutsch equation.

Selection of the value of precipitation rate parameters is dependent upon a number of factors; the most important being the size and resistivity of the dust. The size of the dust from sinter machine wind boxes is relatively large since very little metallurgical fume is generated. Consequently, the precipitation rate parameter would tend to be reasonably high for moderate to low resistivity dust. However, recent trends toward the use of larger quantities of lime in the self-fluxing sinters have resulted in high resistivity dust which reduces power input to the precipitator and makes necessary lower precipitation rate parameters. The addition of the larger amounts of lime changes the nature of the sinter machine dust from one that is relatively easy to one relatively difficult to precipitate. Consequently, it is necessary to define the conditions under which the sinter machine will operate and the characteristics of the dust before the value of the precipitation rate parameter can be selected.

Figure 17.13 is a plot of the efficiency of a limited group of sinter machine precipitators as a function of the plate-area-to-gas-volume-ratio. Design w's are 10-12 cm/sec as shown by the curves. The test data, however, show a much wider spread in values. This probably reflects the variations in the type of dust being collected.

Cost data. Based on a limited sample, the costs of precipitators for sinter machine service ranges from about $10.50 to $14.00/sq ft of collecting surface. In terms of gas volume handled, the erected costs range from around $1.00 to $2.50 per cu ft of gas handled. Figures 17.14 and

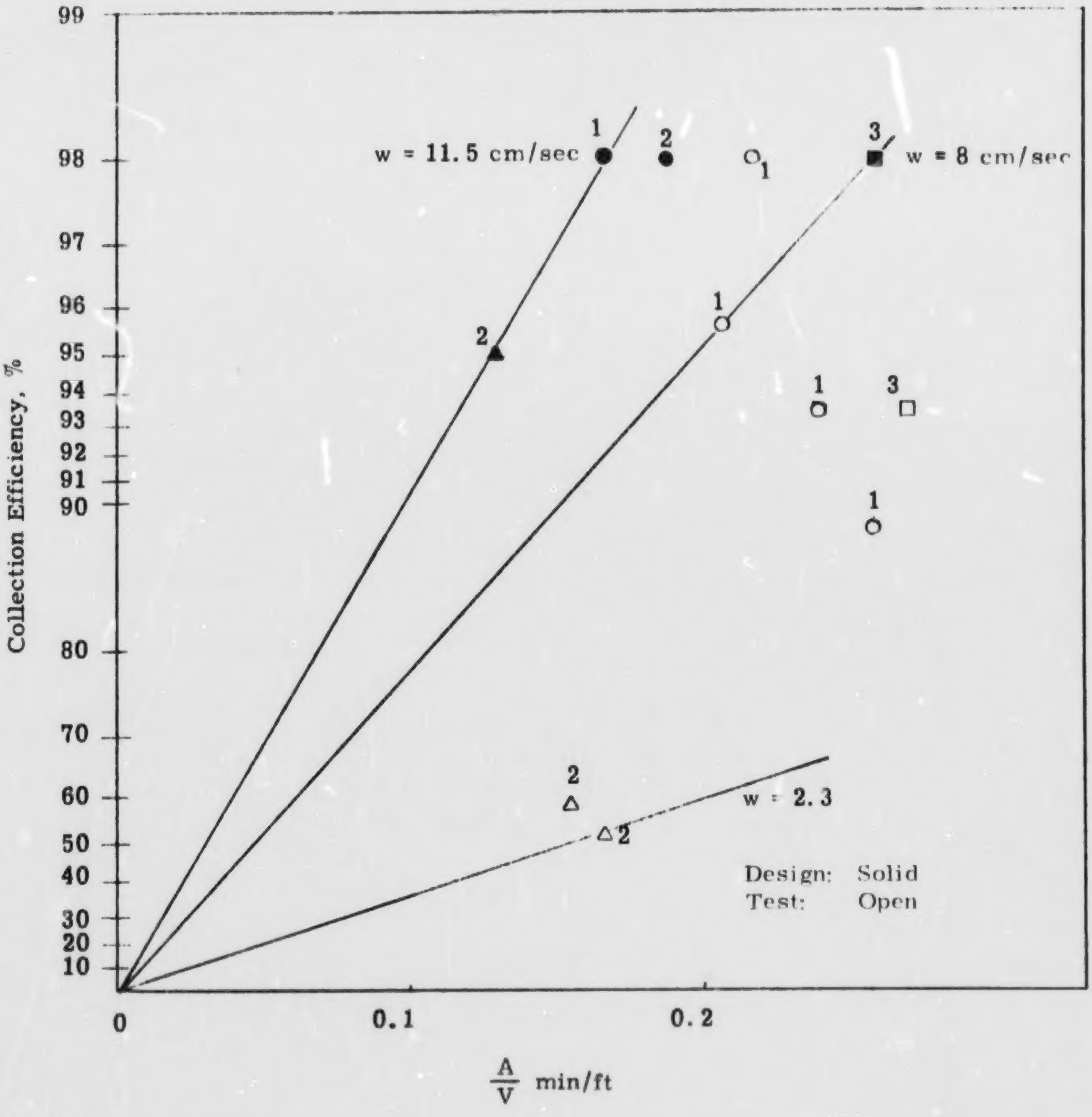


Figure 17.13. Variation in Collection Efficiency of Sinter Machine Precipitators with Collection Surface Area to Gas Flow Ratio.

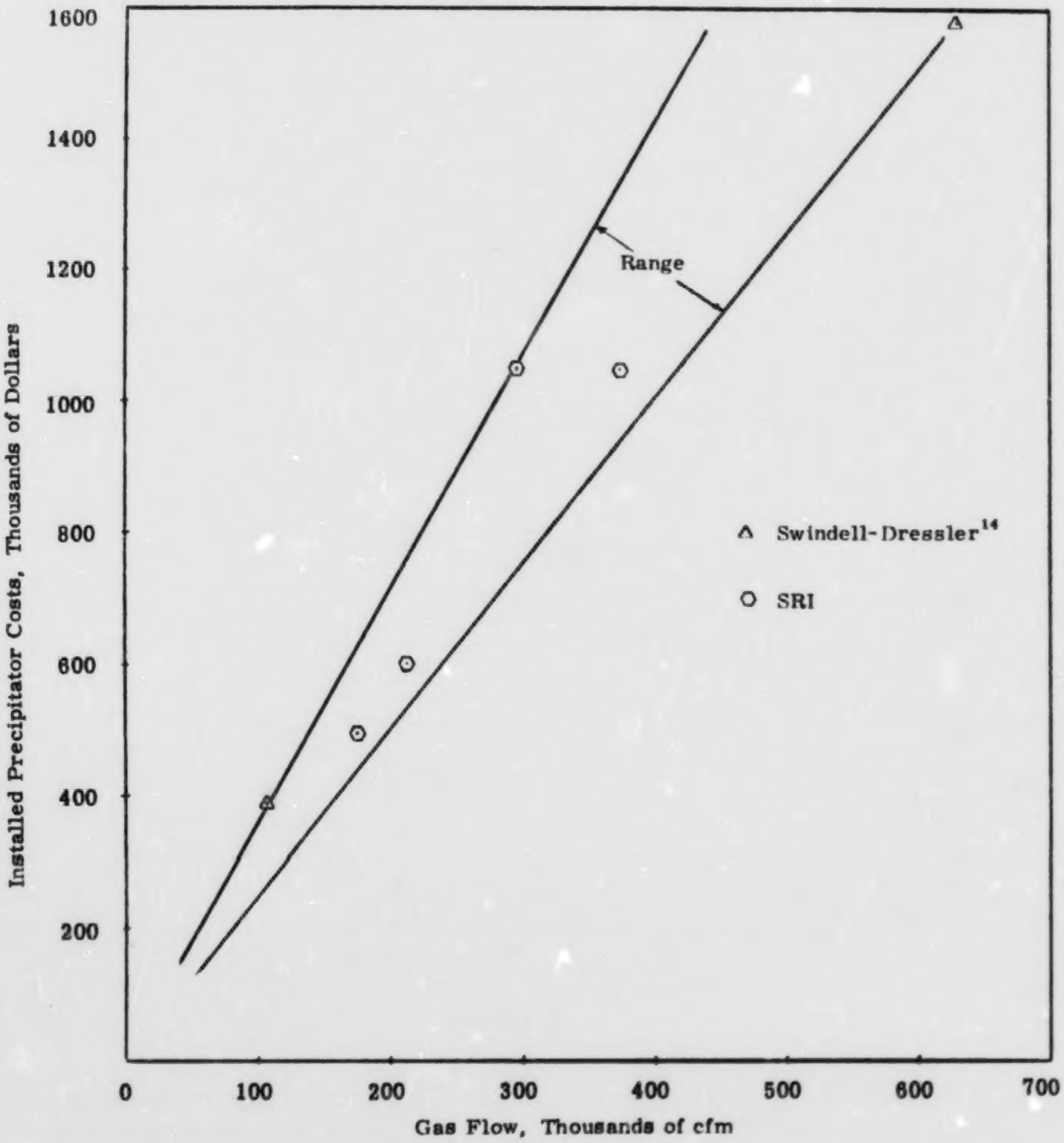


Figure 17.14. Range of Installed Costs For Various Gas Volumes for Precipitators Operating On Sinter Machines. Design Efficiency 98%.

17.15 show the range of flange-to-flange precipitator costs and installed costs for various gas volumes handled. Costs are for the 1967-1969 period and again are based on a limited sample.

Table 17.6 lists capital and operating costs for electrostatic precipitators on sinter machines as prepared by Swindell-Dressler.[14] These data are also plotted on the curve in Figure 17.14, along with the Southern Research Institute data.

17.4 APPLICATION TO BLAST FURNACES

The blast furnace is a cylindrically shaped, refractory lined structure usually about 100 ft tall and 25-35 ft in diameter. A composite sketch of the blast furnace and associated equipment is shown in Figure 17.16. The raw materials used in the production of pig iron are semi-continuously charged into the blast furnace. The charge consists of coke, limestone, and the iron-bearing raw materials. The latter may include screened or unscreened iron ore, sinter oxide pellets, or scrap. Combinations of sinter and oxide pellets comprise the normal iron-bearing charge materials. The raw materials are charged into the top of the furnace through a double bell arrangement which makes it possible to maintain a gas tight seal during charging to prevent atmospheric emissions. This is accomplished by hoisting the raw materials to the top of the furnace and discharging them into a hopper located above the upper bell. When this bell is opened, the charge drops through into an area above the lower bell. The upper bell is then closed and the lower one opened, permitting the charge to fall into the furnace without the escape of dust laden gas.

The furnace is heated, and the reducing atmosphere necessary for reduction of iron ore to iron is generated by a reaction between the precharged coke and preheated air. The air is introduced through tuyeres located in the lower portion of the furnace. The air blast is usually preheated to temperatures of 1000-1700°F. It may be enriched with natural gas, fuel oil, or oxygen. As the blast rises through the burden, it reacts exothermically with the coke to produce the high temperature reducing gases which react with the iron oxide to produce molten pig iron. The molten iron collects in the lowest region of the furnace, known as the hearth. The hearth is usually tapped every 3-4 hours, with a yield of from 100 to 300 tons of hot metal per tap.[15,16]

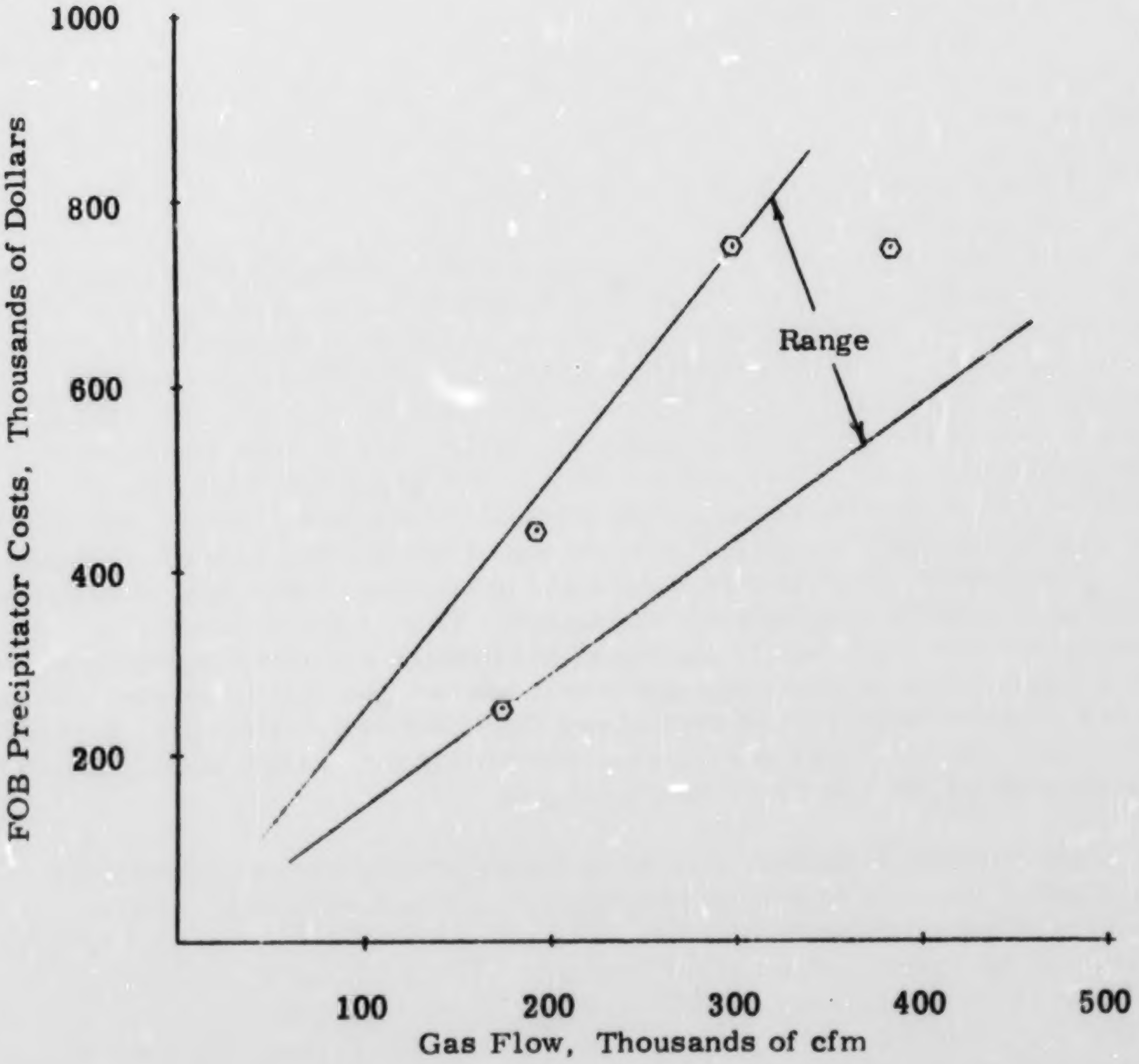


Figure 17.15. Range of FOB Precipitator Costs for Various Gas Volumes. Sinter Machine Installations for 98% Design Efficiency.

Table 17.6

Sinter Plant - Windbox - Electrostatic Precipitator[14]

Capital Costs

Gas Volume - acfm at 325°F	105,000	630,000
Plant Capacity - tons/day	1,000	6,000
1. Material	$ 210,000	$ 955,000
2. Labor	88,000	325,000
3. Central Engineering	73,000	240,000
4. Client Engineering	18,000	60,000
Total	$ 389,000	$ 1,580,000

Operating Costs ($ /Yr)

Gas Volume - acfm at 325°F	105,000	630,000
Plant Capacity - tons/day	1,000	6,000
1. Electric Power	$ 12,500	$ 77,000
2. Maintenance	16,000	60,000
3. Operating Labor	20,000	30,000
Direct Operating Cost	$ 48,500	$ 167,000
4. Depreciation	39,000	158,000
5. Capital Charges	39,000	158,000
Total	$ 126,500	$ 483,000

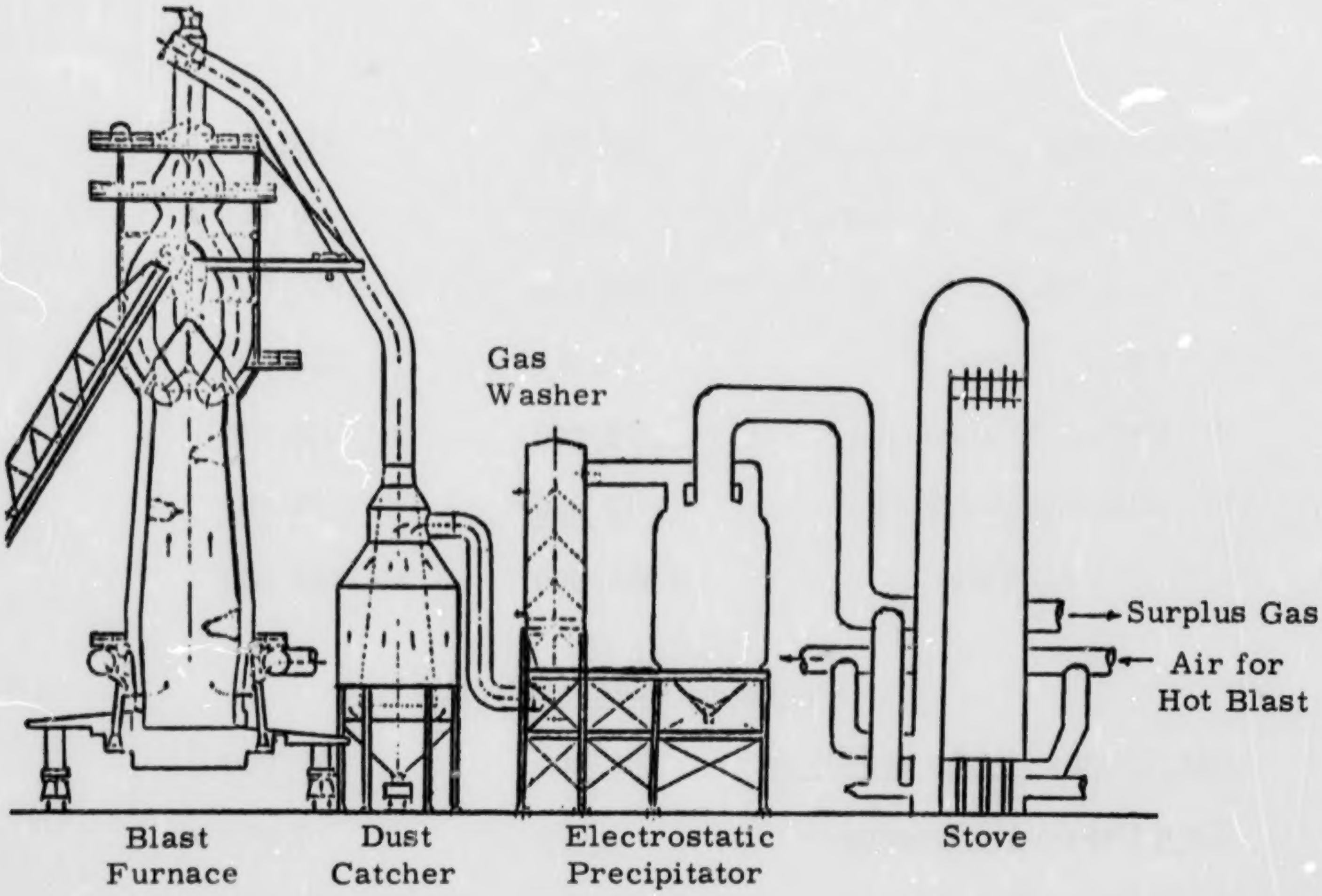

Gas
Washer
Surplus Gas
Air for
Hot Blast
Blast
Furnace
Dust
Catcher
Electrostatic
Precipitator
Stove

Figure 17.16. Flow Diagram for Wet Cleaning Iron Blast Furnace Gas With Electrostatic Precipitator.

After passing through the charge, the gases leave the top of the furnace at a temperature of around 300°F. The composition of the gases is typically as follows.

Constituent	Volume Percentage
CO	26.2%
CO_2	13.0%
H_2	1.9%
CH_4	0.2%
N_2	58.7%

These gases leave the top of the furnace through four "off takes" arranged around the dome, and pass through a brick-lined downcomer to the dustcatcher, which is a large settling chamber. The force of gravity causes the coarser dust particles to settle in the low gas velocity regions of the settling chamber.

The particulate material emitted during normal operation of the blast furnace comes from several sources, including dirt and other fines in the charged ore, the dust developed by the downward abrasive action of the burden through the furnace, and coke and limestone dust.[17] These particulates are carried out of the furnace by the gas stream, with a dust loading in the range of from 7 to 17 gr/scf.

A troublesome source of emissions from the blast furnace is that occurring during a burden "slip." A "slip" occurs when a portion of the burden bridges within the furnace while some of the underlying burden continues to melt and move downward in the furnace. After a gap develops between the two regions, the upper burden may become dislodged and slip. When this occurs there is an almost explosive release of dust-laden gases, and these must be vented directly to the atmosphere for safety reasons, and to avoid the possibility of damage to the furnace.[17,18]

The output of particulate material from blast furnace operations is primarily dependent on the physical characteristics of the charge, although start-up operations may produce some variations in the type and amount. During the start-up period, heavy fuel oil, tar or natural gas may be injected into the furnace along with the heated air. Because of the low blast and coke temperatures, incomplete combustion may take place and give rise to soot production. As the blow-in progresses and as operating temperatures are reached, soot production decreases until, during normal operations, no soot is produced.[2,5]

Sinter and pelletized ore, when used as a charge material, effect a marked reduction in dust product as compared to either screened or unscreened iron ore, because of the improved physical characteristics. This is illustrated in Figure 17.17, where the amount of dust from a furnace is shown as a function of the amount of sinter in the charge. The use of 3100 lb of sinter and taconite pellets charged per net ton of hot metal produced is defined as a completely beneficiated charge. It is to be noted that an increase in wind volume from about 85,000 cfm to almost 115,000 cfm has accompanied the use of beneficiated iron ore. Simultaneously, the dust production per net ton of hot metal has decreased by a factor of five.[17]

Composition of particulate and gas emissions. During normal blast furnace operations, the dust loading of the gas leaving the furnace is in the 7-10 gr/scf range[17] although top gas dust loadings as high as 17 gr/scf have been reported.[5] For each ton of hot metal produced, 110,000-150,000 cu ft of waste gas is produced, and this carries from 50-300 lb of dust from the furnace.[2,4,17] For each 1000 tons of hot metal produced, about 100 tons of particulate is expected. Approximately 68% of this will be over 50μ in diameter.[19]

The chemical composition of the particulate varies considerably as shown in Table 17.7. The major constituents and the usual composition range is as follows: 35-50% iron, 8-13% silica, 2-5% alumina, 3-4% calcium oxide, 3-10% carbon and small amounts of alkali elements, phosphorus, zinc, lead, sulfur, and other trace elements.

Some additional data on the chemical composition of dust samples taken from the dust catchers, wet scrubbers, electrostatic precipitators, and dust leaving the electrostatic precipitator are given in Table 17.8. The material collected in flue dust catchers and washers has a high iron content, and after agglomeration, makes a suitable material to be charged back into the furnace. The dust from the precipitator has a lower iron content and contains several objectionable impurities.

The particle size distribution of blast furnace dust may vary between wide limits, depending on the type of ore being charged into the furnace and prior beneficiation. Table 17.9 presents a range of particle sizes that have been observed.

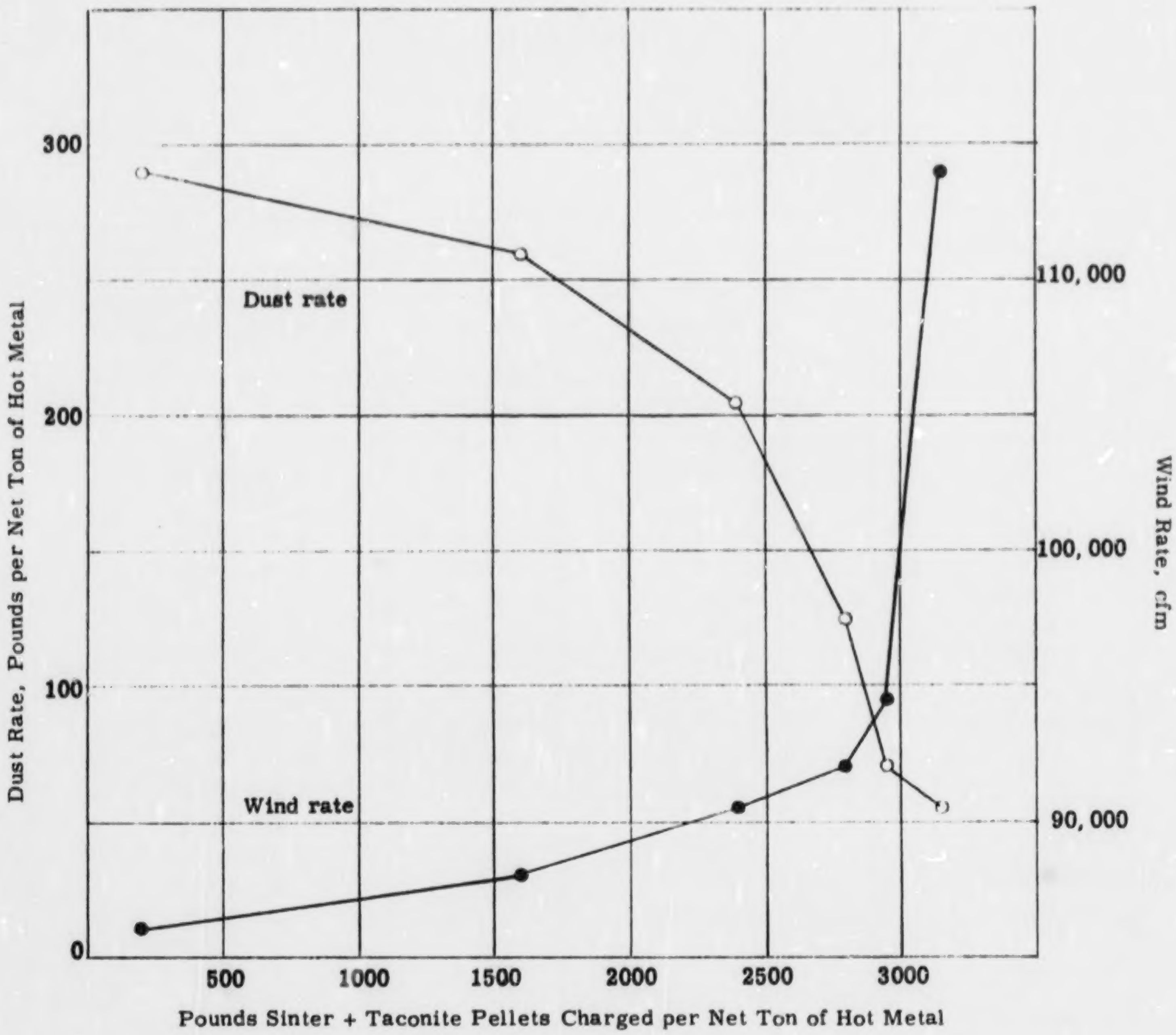


Figure 17.17. Dust Rate and Wind Rate as a Function of the Sinter Plus Taconite in the Blast Furnace Burden.

Table 17.7

Chemical Analysis of Blast Furnace Flue Dusts

Components	Weight Percent		
	Range for Several U. S. Plants	Midwest Plant	Range for Several European Plants
Fe	36.5 - 50.3	47.10	5.0 - 40.0
FeO	n. a.	11.87	n. a.
SiO_2	8.9 - 13.4	8.17	9.0 - 30.0
Al_2O_3	2.2 - 5.3	1.88	4.0 - 15.0
MgO	0.9 - 1.6	0.22	1.0 - 5.0
CaO	3.8 - 4.5	4.10	7.0 - 28.0
Na_2O	n. a.	0.24	-
K_2O	n. a.	1.01	-
ZnO	n. a.	0.60	-
P	0.1 - 0.2	0.03	0.3 - 1.2
S	0.2 - 0.4	n. a.	- 0.1
Mn	0.5 - 0.9	0.70	0.3 - 1.5
C	3.7 - 13.9	n. a.	5.0 - 10.0
Pb	n. a.	n. a.	0 - 15.0
Cu	n. a.	n. a.	trace
Alkali	-	-	0 - 20.0
Zn	n. a.	n. a.	0 - 35.0

Table 17.8

Weight Percent Composition of Dust Samples from Blast Furnace Gas Cleaning Plant

Sample Components	Deposit from primary dust catcher	Deposit from secondary dust catcher	Deposit from washers	Dust in gas leaving washers	Deposit from precipitator	Dust in gas leaving precipitator	Deposit in power station burners
Insol.	8.50	11.60	9.40	13.68	22.60	11.44	
Fe_2O_3	75.43	53.00	61.30	14.40	19.30	8.64	3.52
Al_2O_3	1.80	1.55	4.35	6.91	15.58	6.70	
MnO	0.62	0.60	0.63	2.16	1.10	1.49	
CaO	1.60	2.40	3.56	6.36	7.12	4.68	2.80
MgO	0.80	1.09	1.67	8.04	9.90	9.28	3.04
P	0.24	0.20	0.41	0.54	0.71	0.57	0.14
S	-	-	0.41	1.89	1.33	1.37	0.36
Cl	trace	trace	nil	1.32	0.20	3.44	29.22
Zn	nil	nil	nil	1.20	0.90	1.20	0.70
Na_2O	0.28	0.32	0.25	9.25	1.55	9.75	5.80
K_2O	0.56	0.92	0.55	20.90	2.70	22.40	36.15
Loss on ignition	5.90	23.50	15.65	-	15.95	-	9.06

Table 17.9

Size Analysis of Blast Furnace Flue Dust

Screen Size		Dust Loading
Mesh	Microns	Range, %
20	833	2.5 - 20.2
30	589	3.9 - 10.6
40	414	7.0 - 11.7
50	295	10.7 - 12.4
70	208	10.0 - 15.0
100	147	10.2 - 16.8
140	104	7.7 - 12.5
200	74	5.3 - 8.8
-200	-74	15.4 - 22.6

Cleaning of blast furnace gas. Blast furnace gas has a heating value of about 100 Btu/scf or about one-tenth that of natural gas; thus it is a valuable fuel for heating the blast furnace stoves, steam boilers, soaking pits, and coke batteries. However, the gas must be cleaned prior to burning in order to prevent the clogging of burners, gas mains, and in general, to avoid maintenance problems. In a typical installation, gas from the furnace is generally cleaned as follows. Upon leaving the furnace, it is passed through a dust catcher where the heavier particles are separated by inertial forces. This generally removes particles larger than a few hundred microns (50-70% by weight) and leaves a dust concentration in the gas of 3-6 gr/scf.[4,17] From the dust catcher, the gas is further cleaned in a two-stage plant. The first stage consists of a primary cleaner to separate the coarse fractions and a secondary cleaner to remove fine dust. Many combinations of two-stage cleaning plants can be used, but a typical one consists of a wet scrubber and an electrostatic precipitator in series.[17]

Typical dust concentration for the gas at various points in a modern gas-cleaning system are:

at top of furnace - 7 - 13 gr/acf

dust catcher outlet - 3 - 6 gr/acf

after primary washer - 0.05 - 0.7 gr/acf and,

after electrostatic precipitator - 0.005 - 0.01 gr/acf.

After cleaning, part of the gas produced is utilized in the air blast preheat stoves. The remainder may be used for underfiring boilers or coke ovens, or for other miscellaneous heating applications in the steel plant. Figure 17.16 is a flow diagram showing a typical gas cleaning system for iron blast furnaces.

Blast furnace precipitators. Since the gas delivered to a blast furnace precipitator has been cooled to saturation, a wet-type precipitator is used. The type most widely used is a vertical flow employing pipes of 8-inch or 12-inch diameter as collecting electrodes. Plate-type collecting electrodes with horizontal flow have also been used. Both types are single stage. The shells are constructed of steel and are normally circular for pipes and rectangular for plates.

In some designs the pipes are suspended from a header sheet at the top, with the lower end unsupported, allowing the gas to surround the pipes. In other designs, a lower header sheet is used in order to prevent the gas from surrounding the pipes. Figure 17.18 shows a typical wet pipe-type blast furnace precipitator.

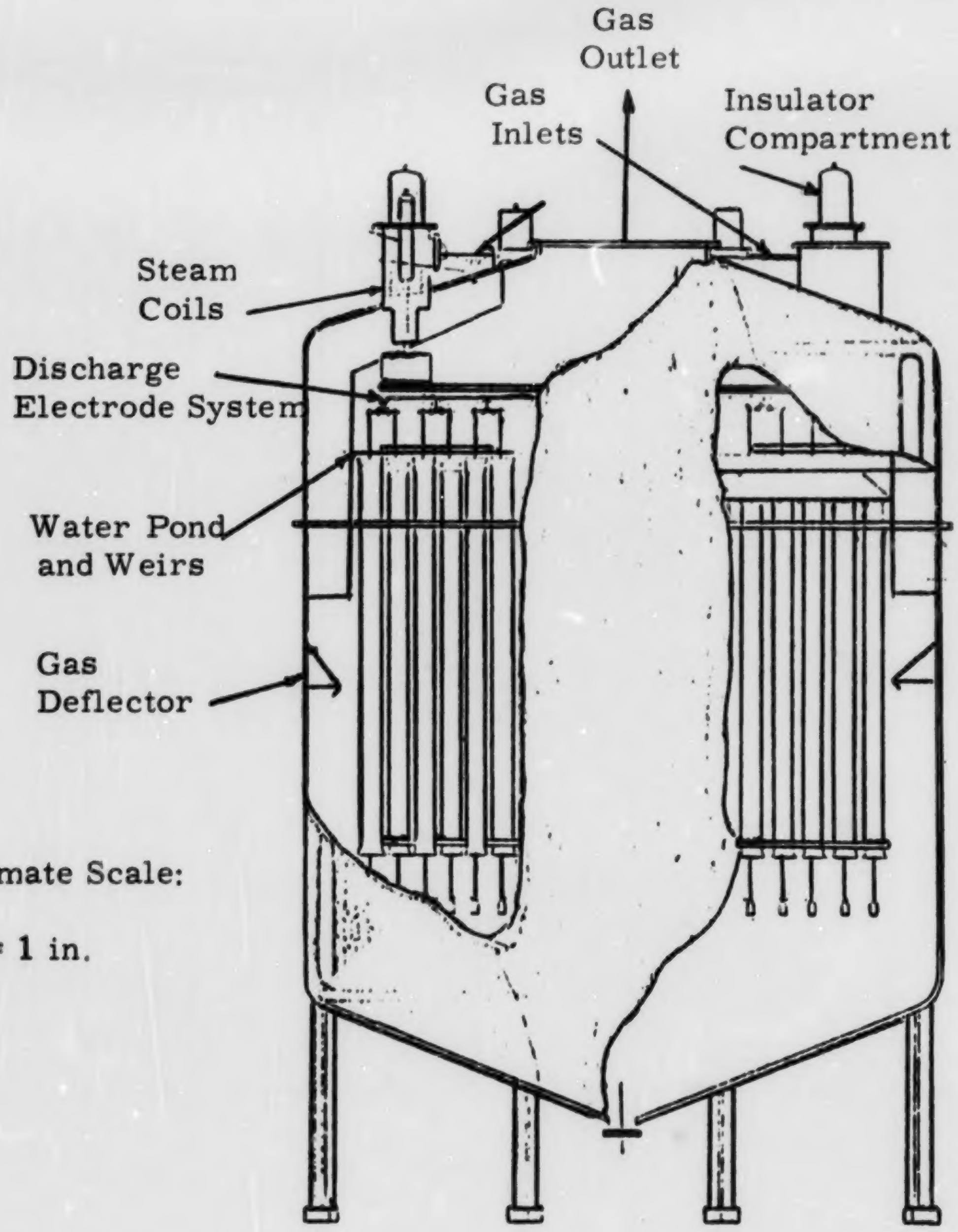
Discharge electrodes are usually vertically hung, twisted square bars (3/16 - 1/4 in.) spanning the height of the collecting electrodes. Electrode rapping is not required since the collectors are flushed with liquid.

The particulate removal system for both vertical and horizontal type precipitators is through a slurry hopper. In most cases there is enough flushing and/or spray water used to permit slurry removal from the hopper by simple draining. In a few cases where the collected particulate becomes bulky, or where the slurry becomes thick, an agitator or stirring system is installed. The slurry is normally piped to a sewer or to storage.

Precipitator installations. Trends in the application of electrostatic precipitators for cleaning blast furnace gas are indicated by the number of installations from 1931 to 1968 given in Table 17.10. Figure 17.19 shows the number of furnaces and the number of furnaces with electrostatic precipitators installed. Figure 17.20 shows the precipitator installations over the period 1931-1969 on the basis of the gas volume treated.

Precipitator application and design data. Precipitator inlet gas temperatures and inlet particulate loading are shown in Figures 17.21 and 17.22. Distribution of Precipitator gas velocity, field strength and power are shown in Figure 17.23 - 17.25. Design efficiency trends are shown in Figure 17.26.

Design values of precipitation rate parameter w for pipe-type blast furnace precipitators are usually in the range of 0.25 to 0.45 ft/sec. Performance values of w tend to be somewhat higher, typically from 0.30 to 0.60 ft/sec. There is statistical evidence that the value of w increases with the gas velocity through the precipitator, possibly owing to improved gas flow quality at the higher gas velocities. It is therefore often advantageous to design for relatively high gas velocities, since this means that higher values of w can be used and the total amount of collection surface required can be correspondingly reduced. Note that the wet-film collection surface serves to prevent reentrainment of collected particles by aerodynamic forces. Therefore, much higher gas velocities can be used


Gas Outlet
Gas Inlets
Insulator Compartment
Steam Coils
Discharge Electrode System
Water Pond and Weirs
Gas Deflector

Approximate Scale:

5 ft = 1 in.

Figure 17.18. Typical Wet Type Pipe Precipitator for Cleaning Blast Furnace Gases.

Table 17.10

Distribution of Total and Electrostatically Controlled Gas Volume of Effluent from the Blast Furnace from 1931 to Present

Pptr. Contract Year	No. of Install.	No. of Pptrs.	Total Gas Vol. 10^6 acfm	Total Gas Vol. Accumul. 10^6 acfm	Five (5) Year Period: Total Gas Vol. (10^6 acfm)	Five (5) Year Period: Avg. Vol./Yr. During Period (10^6 acfm)	Weighted Design Efficiency on acfm Basis: Yearly	Weighted Design Efficiency on acfm Basis: 5-Yr. Period
1931	3	3	.079	.079	.1474	.0295	95.4	92.6
1933	1	1	.0684	.1474			90.0	
1936	9	23	.5665	.7139	1.063	.2126	92.4	93.1
1937	6	13	.4015	1.1154			93.2	
1938	1	2	.066	1.1814			98.0	
1939	1	1	.029	1.2104			95.0	
1940	6	6	.3825	1.5929	2.472	.4945	92.2	91.9
1941	9	12	.6581	2.251			92.8	
1942	15	21	1.3538	3.605			91.3	
1943	2	2	.078	3.683			91.6	
1945	1	1	.117	3.800	2.194	.439	90.0	91.4
1946	5	7	.649	4.449			90.0	
1947	10	11	1.032	5.481			91.9	
1948	3	4	.244	5.725			91.7	
1949	2	3	.152	5.877			95.0	
1950	4	4	.395	6.272	3.086	.6172	95.7	94.7
1951	16	23	1.337	7.609			95.6	
1952	3	4	.245	7.854			93.0	
1953	7	16	.928	8.782			93.2	
1954	3	4	.181	8.963			94.5	
1955	6	6	.274	9.237	1.963	.393	92.3	93.2
1956	11	13	1.157	10.394			94.0	
1957	4	8	.479	10.873			91.7	
1959	1	1	.053	10.926			95.0	
1960	2	2	.167	11.093	.375	.075	91.5	93.7
1961	1	1	.105	11.198			95.0	
1962	1	1	.103	11.301			96.0	
1968	1	1	.060	11.361	.060	.012	98.0	98.0
Grand Totals	136	196						

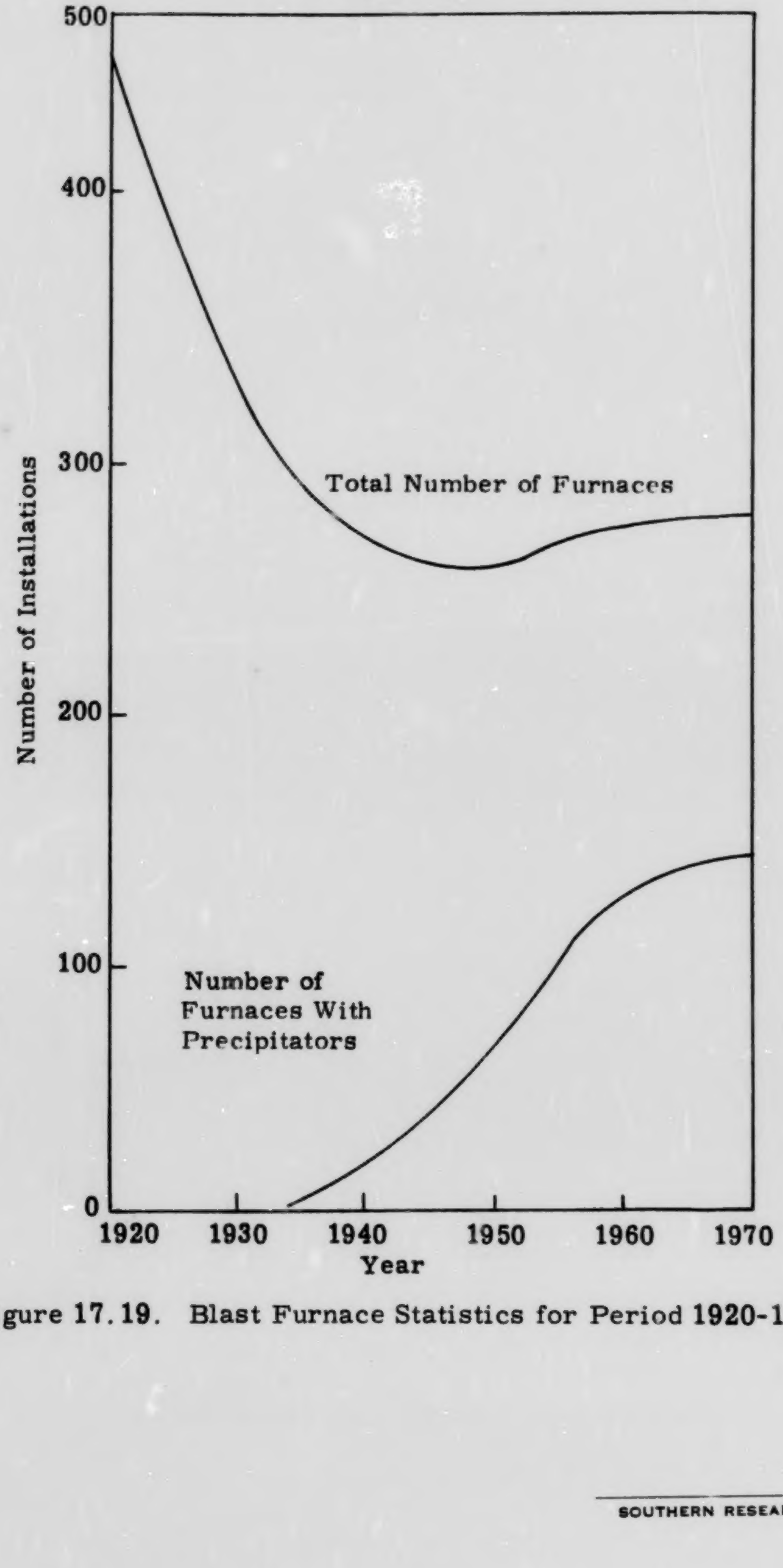


Figure 17.19. Blast Furnace Statistics for Period 1920-1969.

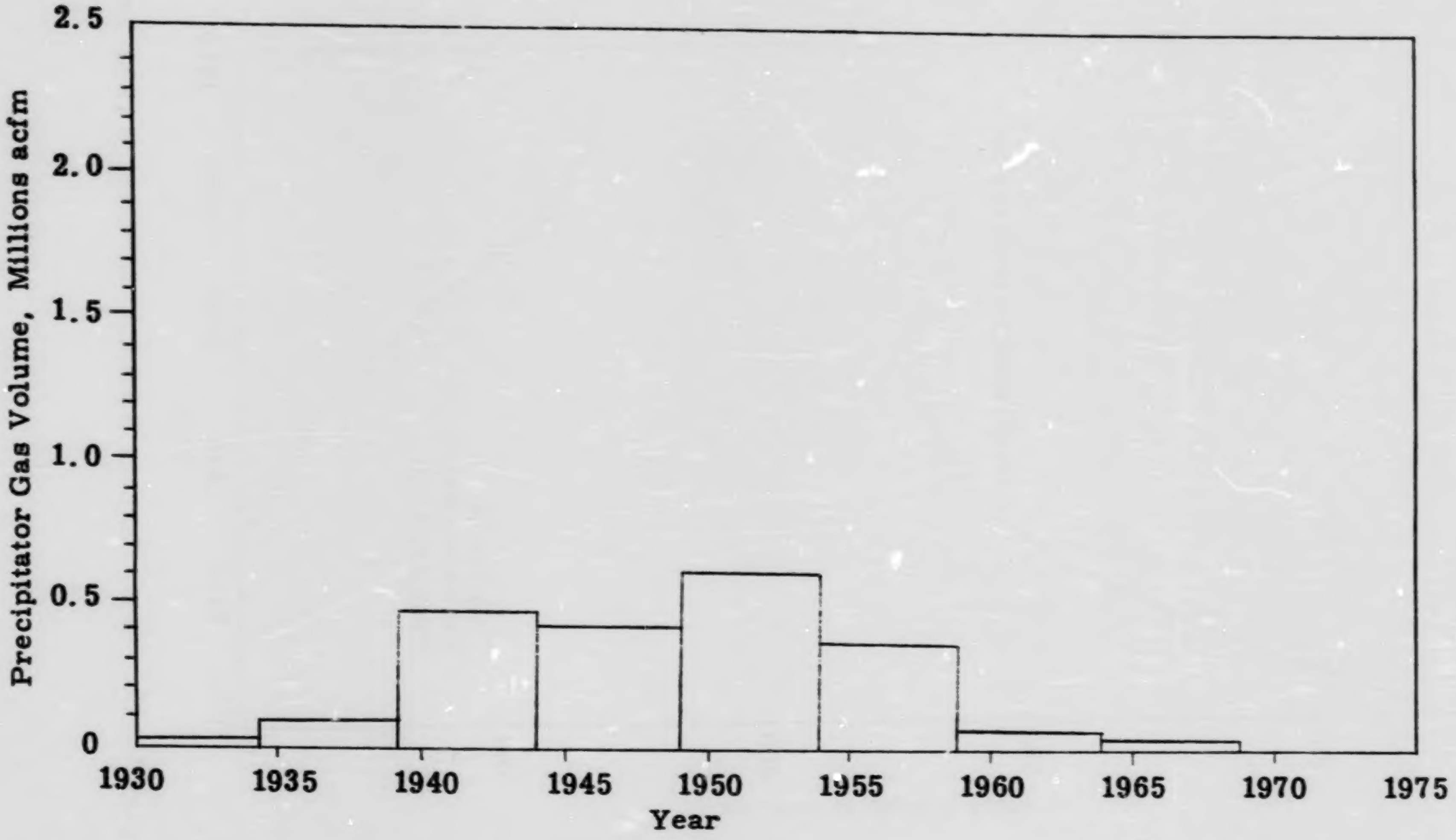


Figure 17.20. Blast Furnace Precipitator Gas Flow Capacity for the Period 1931-1969.

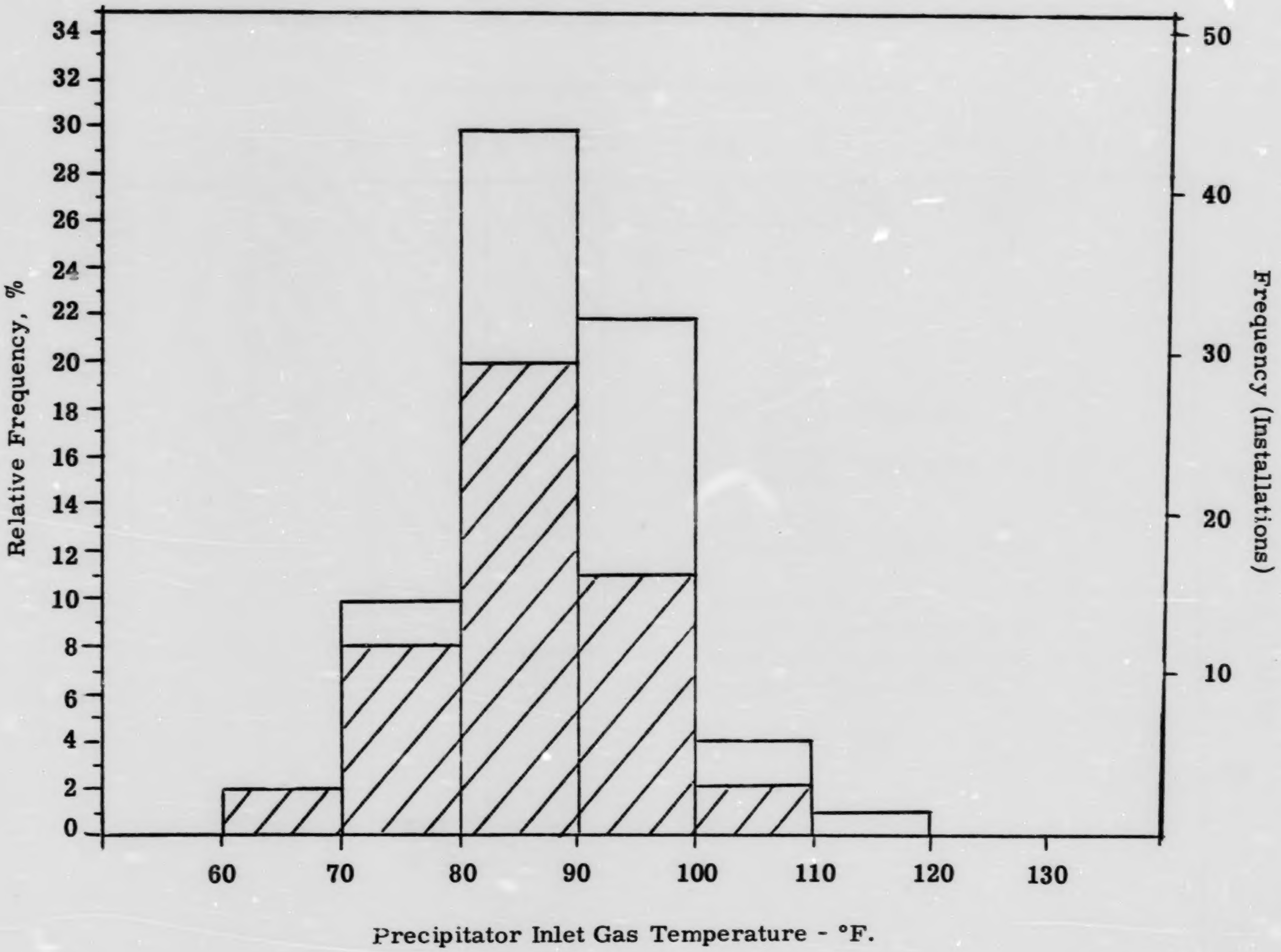


Figure 17.21. Distribution of Precipitator Inlet Gas Temperature. Blast Furnace Installations.

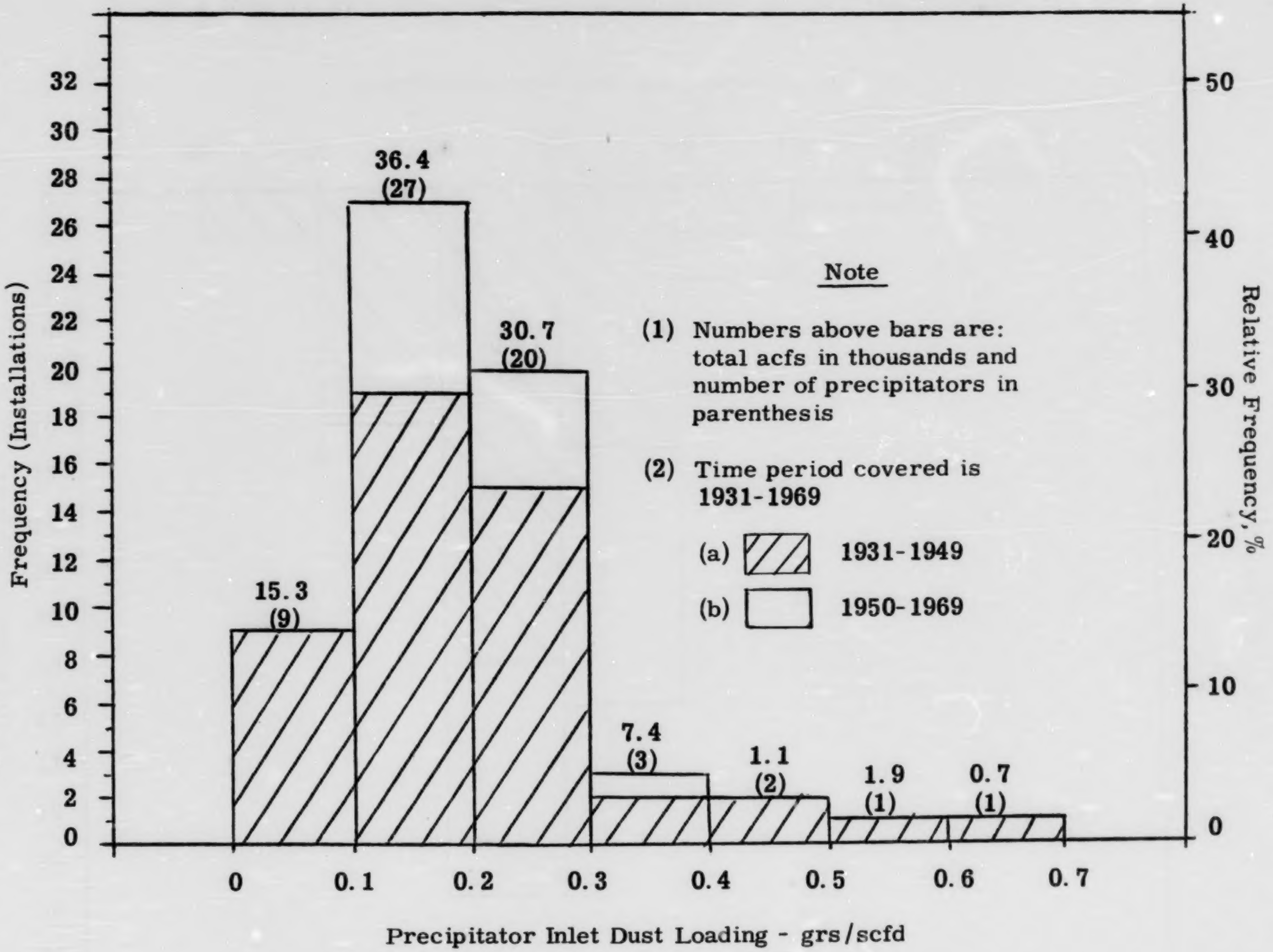


Figure 17.22. Distribution of Precipitator Inlet Dust Loading. Blast Furnace Installations.

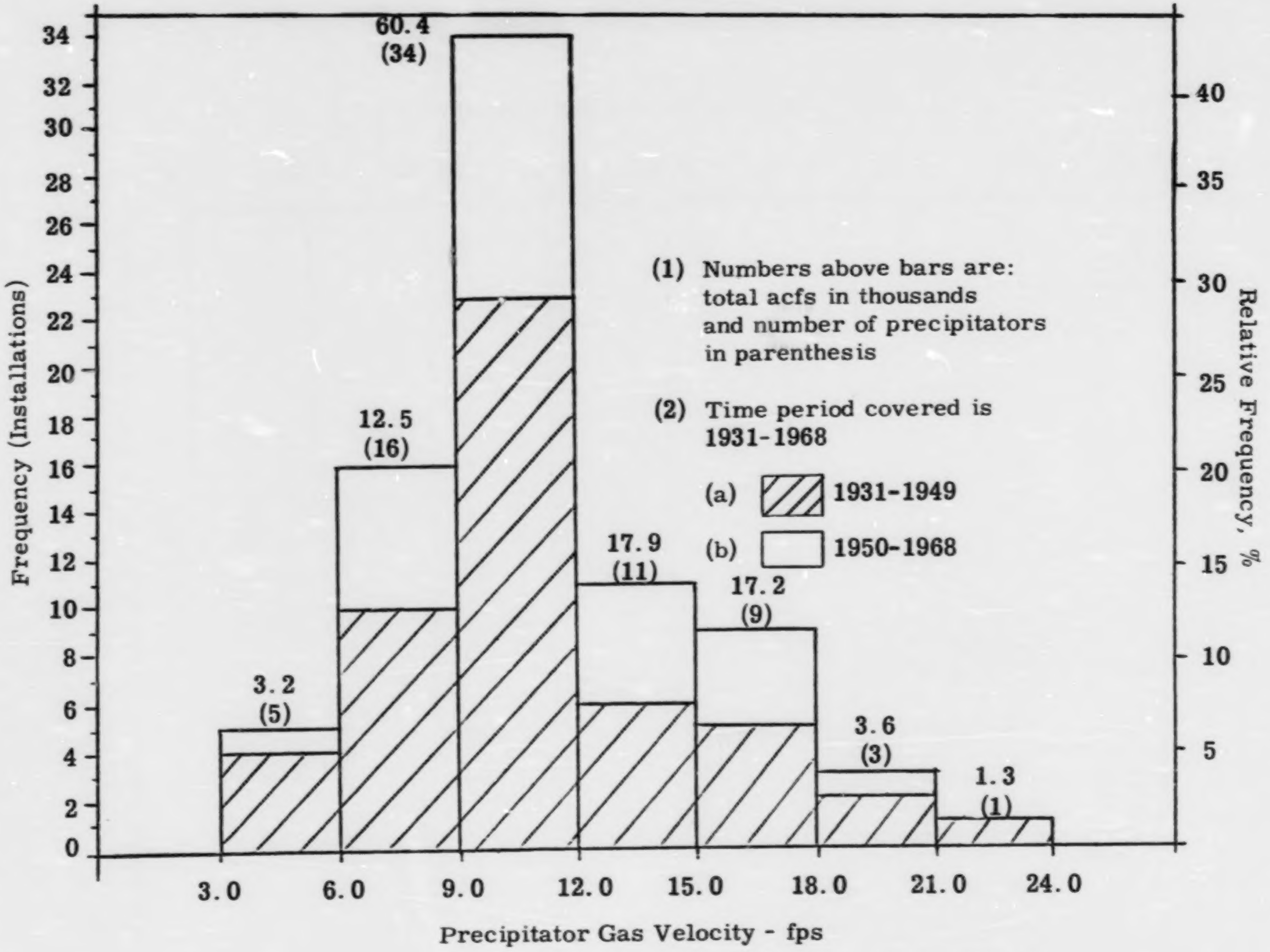


Figure 17.23. Distribution of Precipitator Gas Velocity. Blast Furnace Installations.

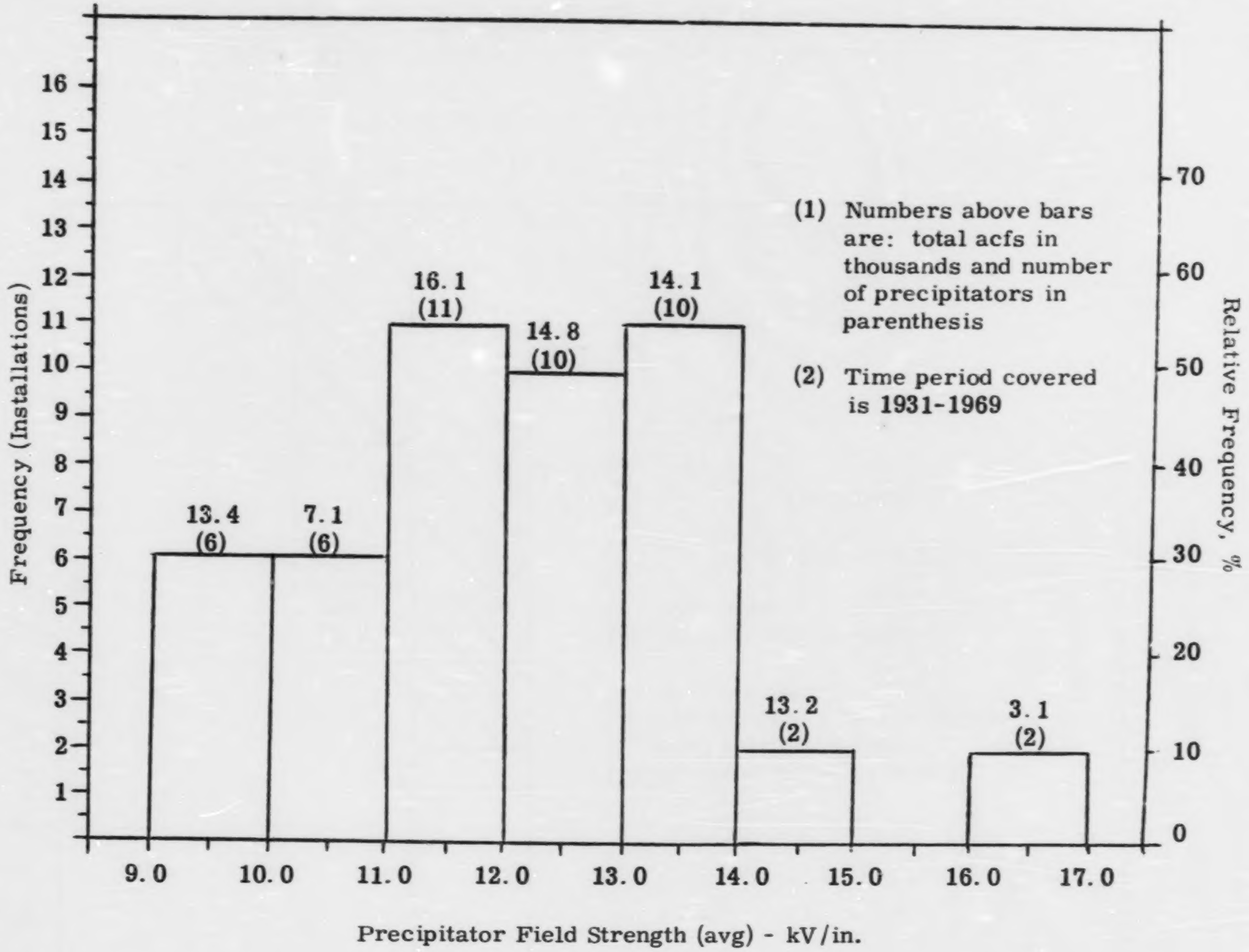


Figure 17.24. Distribution of Precipitator Field Strength.
Blast Furnace Installations.

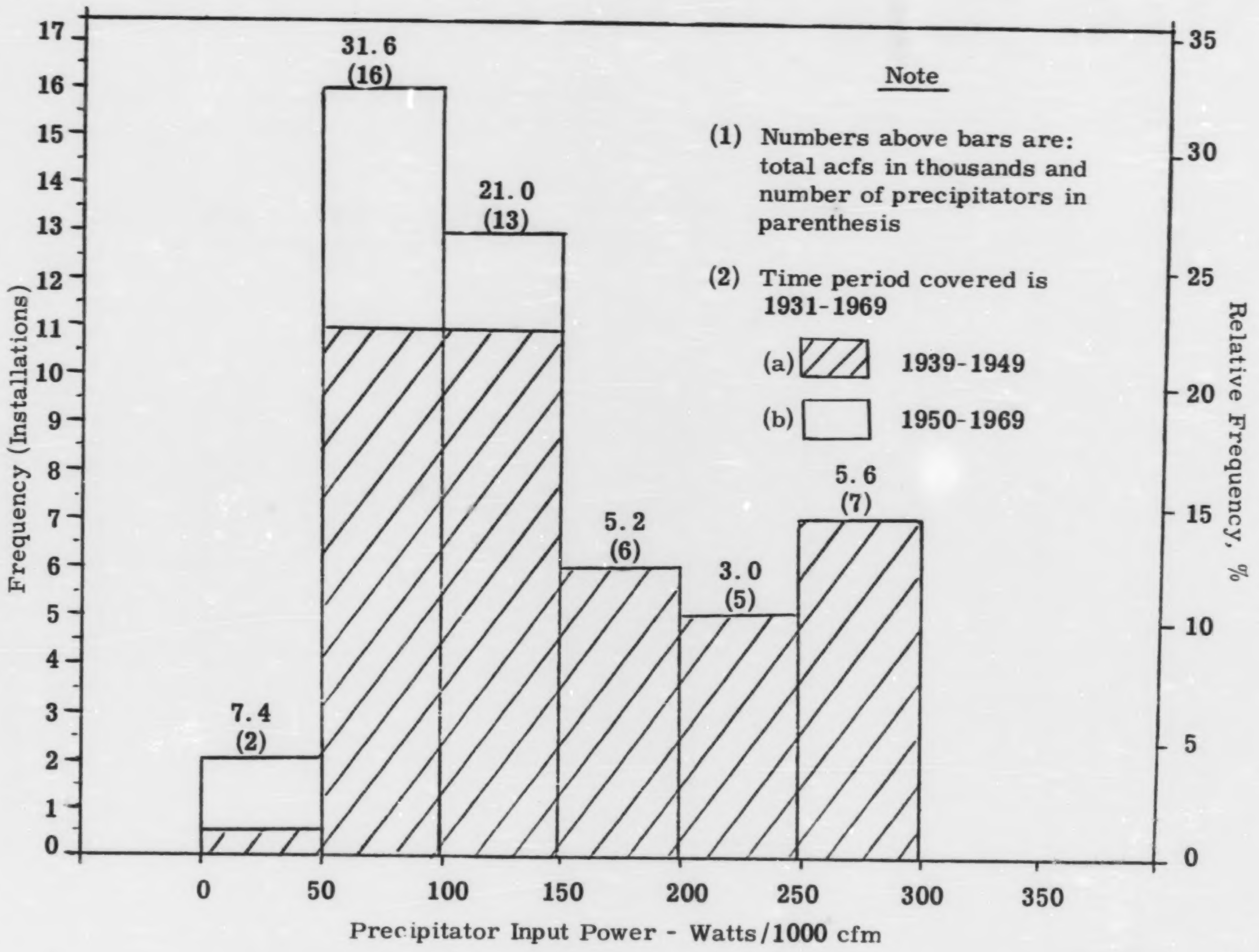


Figure 17.25. Distribution of Precipitator Input Power. Blast Furnace Installations.

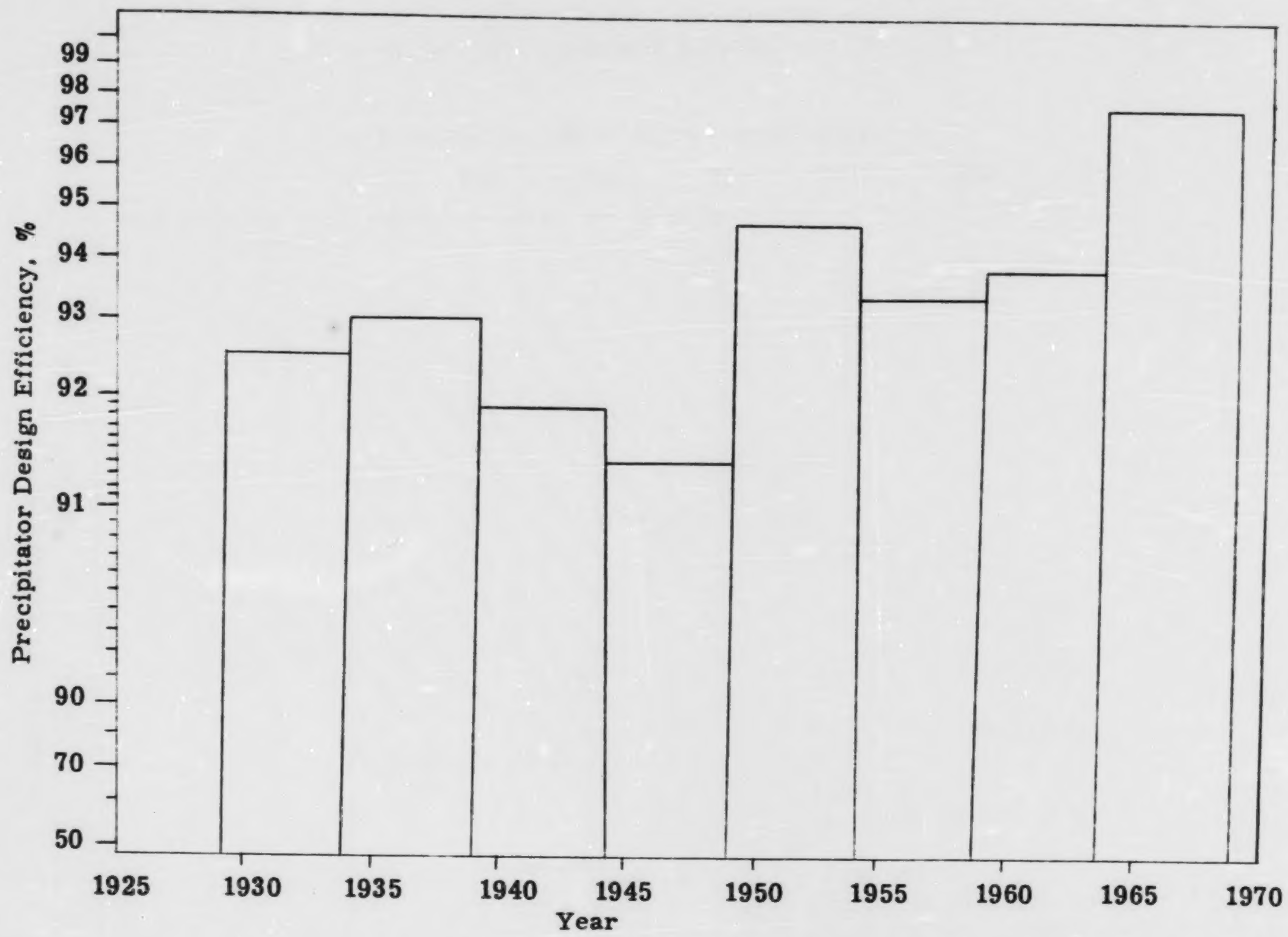


Figure 17. 26. Design Efficiency Trends Over the Period 1931-1969 Prorated on acfm Basis for Blast Furnace Installations.

in the precipitator than is permissible with dry collection.

Figure 17.27 is a plot of the efficiency of blast furnace precipitators as a function of the collection plate area to gas volume ratio. The range of design values is given by the curves, and test data are plotted as individual points. The design data would give precipitation rate parameters ranging from about 9-11 cm/sec (0.3-0.37 ft/sec. Test data show a fairly wide scatter band. The following example will serve to illustrate the design parameters.

Performance Data Based On An
Example Given by Pier[20]

Precipitator - 492 pipes, each 8" dia x 15' long

$$A = \text{collection surface} = 492 \times \pi \times \frac{8}{12} \times 15 = 15{,}400 \text{ sq ft}$$

$$\frac{A}{v_g} = \frac{15{,}400}{160{,}000/60} = 5.78 \text{ sec ft}$$

Inlet conc. = 0.227 gr/cf
Outlet conc. = 0.0147 gr/cf

$$\eta = \text{efficiency} = 1 - \frac{0.0147}{0.227} = 93.5\%$$

$$w = \frac{v_g}{A} \ln \frac{1}{Q} = \frac{1}{5.78} \ln \frac{1}{0.065} = \frac{1}{5.78} \ln 15.4 = \frac{2.73}{5.78} = 0.47 \text{ ft/sec}$$

$$v = \text{gas velocity} = \frac{v_g}{A_c} = \frac{160{,}000}{60 \times 492 \times \pi (1/3)^2} = 15.5 \text{ ft/sec}$$

Cost data. Cost data available for blast furnace precipitators cover a wide range of efficiencies. This coupled with the limited number of installations on which cost data are available and the age of some of the installations, precludes the derivation of statistically meaningful data on which to base cost as a function of gas volume. FOB costs vary from $0.35 to $1.80 per cfm. Erected costs range from $0.70 to $2.40 per cfm.

On the basis of cost per sq ft of collecting surface area, the data are somewhat more meaningful, as indicated in Figure 17.28. Those costs are heavily weighted to the 1950-1960 period, and have not been corrected to a common cost index.

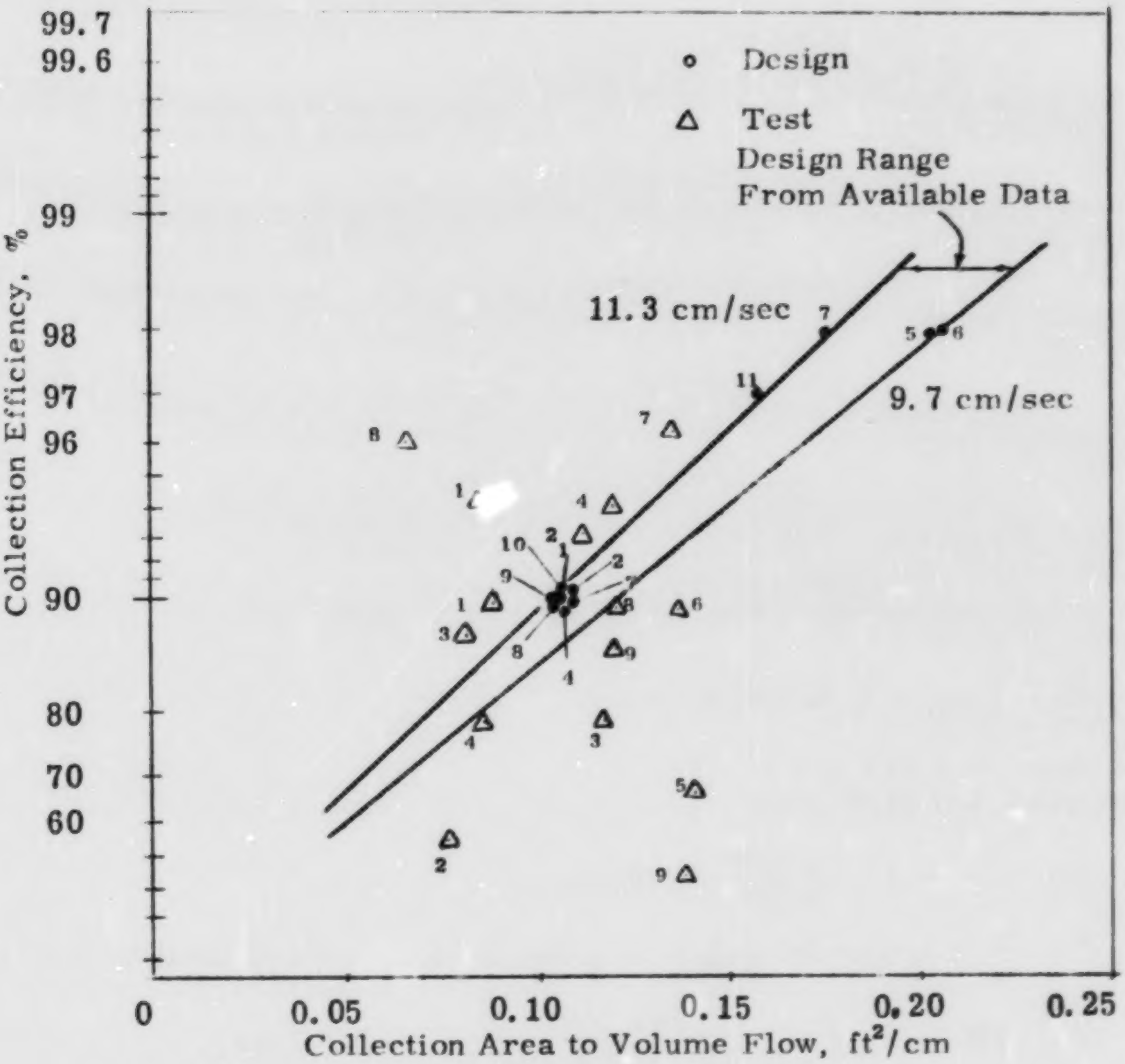


Figure 17.27. Relationship Between Collection Efficiency and Specific Collection Area for Electrostatic Precipitators Operating on Blast Furnace Installations.

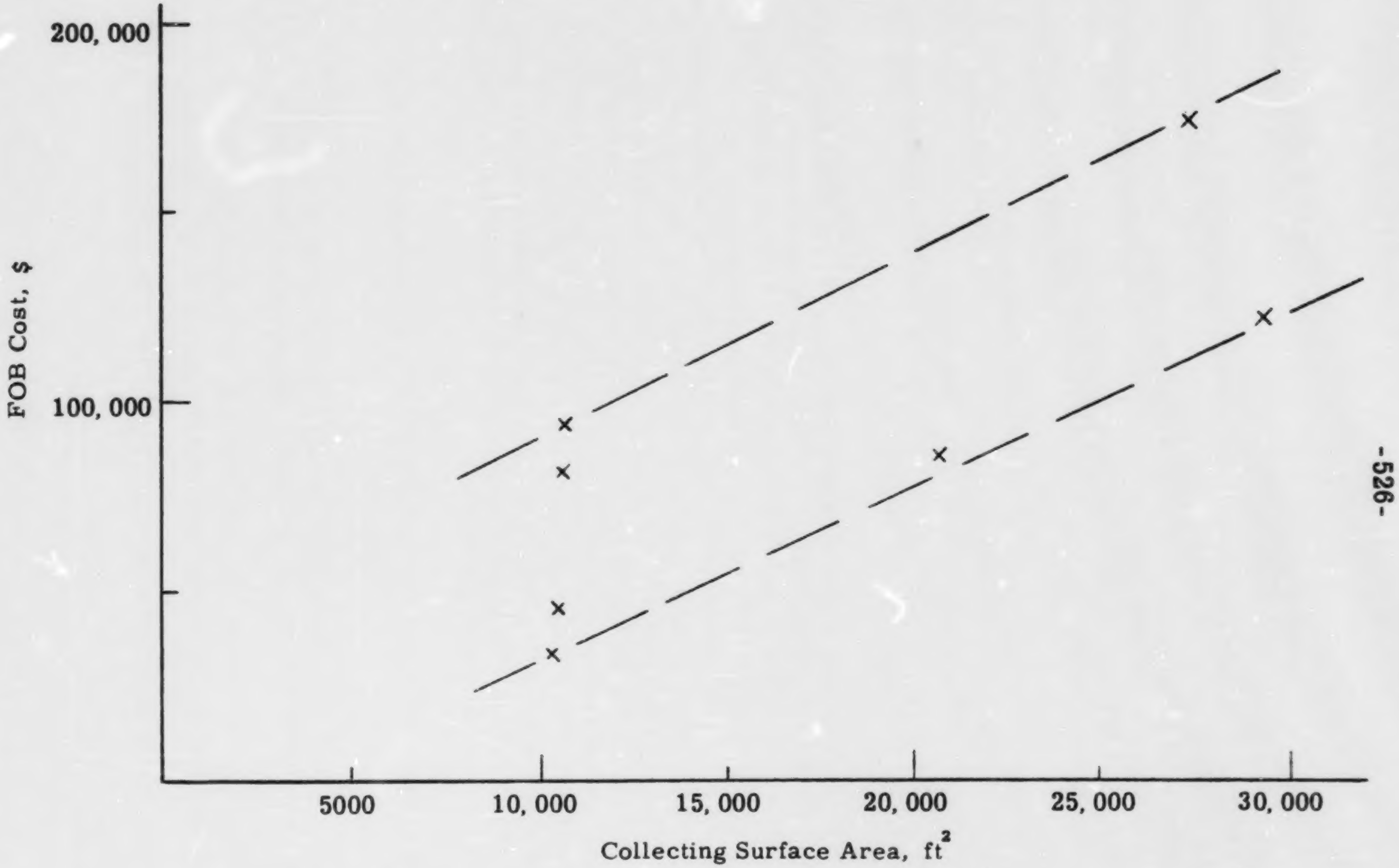


Figure 17. 28. Range of FOB Costs as a Function of Collecting Surface Area for Blast Furnace Installations.

Blast furnaces for ferromanganese production. Another closely allied application to the basic iron blast furnace is one producing ferromanganese. However, this is a more difficult gas cleaning problem since the total solids leaving the dry dust catcher is high, averaging 7-9 gr/scf, and contains a high percentage of fume in the 0.1 to 1.0 micron range. Wet methods of cleaning these gases present difficult problems. When wet, the material forms a hard cement-like deposit which is probably due in part to the fact that the fume is high in alkali content (8-15%).

Alkali originates with the manganese ore, and is concentrated in the escaping particulate by a process of vaporization and condensation. Treatment of scrubbing water, which is not amenable to conventional settling and thickening methods, and stream pollution are other problems encountered with wet cleaning methods.

Figure 17.29 shows a gas cleaning flow diagram for this application using a dry type precipitator following a conditioning tower where the gas temperature is reduced to between 350 and 450°F. Because of its unique chemical and physical properties, the collected dust requires special handling and processing equipment. The method devised involves chemical elements, and pelletizing the material to facilitate handling and storage. Dust flow channels are provided for each precipitator and conditioning tower hopper, to assist in moving the dust to the handling system since continuous removal is essential to avoid packing. The dust is charged to pyrophoring kilns where it is oxidized before going to the mixers and briquetting machines. The production of ferromanganese is rather limited. However, the following illustrative example taken from data given by Pier[20] will indicate the design parameters.

Precipitator treats the combined gases from two ferromanganese blast furnaces.

v = gas flow = 300,000 cfm at 350° to 400°F and about 20% moisture
There are 5 precipitators used in parallel. Each precipitator has 4 sections in series:

$\left\{\begin{array}{l}\text{Inlet section, 23 ducts, each } 10\ 3/4'' \times 4'\text{-}6'' \text{ long} \times 20' \text{ high}\\ \text{Other 3 sections, 27 ducts, each } 8\ 3/4'' \times 4'\text{-}6'' \text{ long} \times 20' \text{ high}\end{array}\right.$

A = collection surface = $5[2 \times 23 \times 4.5 \times 20 + 3(2 \times 27 \times 4.5 \times 20)]$ = 93,600 sq ft

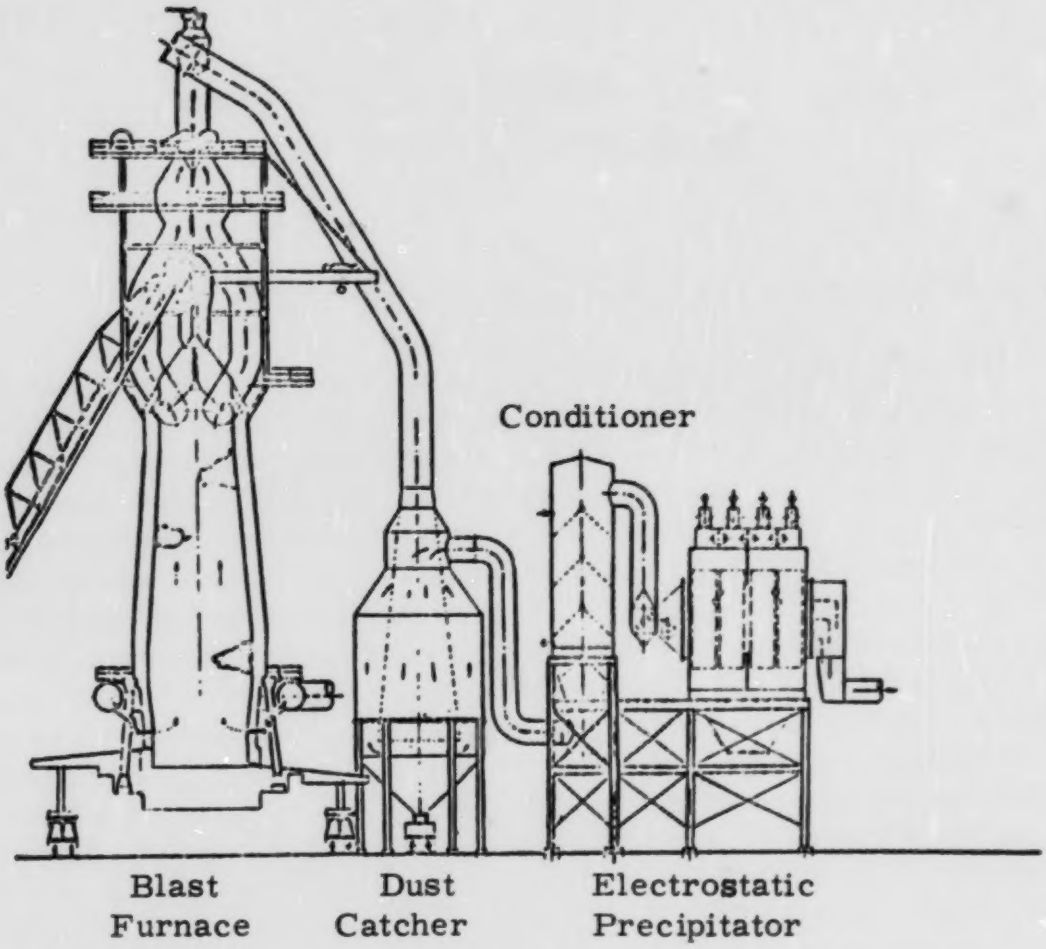

Conditioner
Blast Furnace
Dust Catcher
Electrostatic Precipitator

Figure 17.29. Flow Diagram for Dry Cleaning Ferromanganese Blast Furnace Gas With Electrostatic Precipitator.

$$\frac{A}{v_g} = \frac{93,600}{300,000/60} = 18.7 \text{ sec ft}$$

Inlet conc. $\cong$ 9.0 gr/cf (NTP), outlet conc. $\cong$ 0.15 gr/cf (NTP)

$$\text{Efficiency} = 1 - \frac{0.15}{9.0} \cong 98.3\%$$

$$w = \frac{v_g}{A} \ln \frac{1}{Q} = \frac{1}{18.7} \times \ln \frac{1}{0.017} = \frac{1}{18.7} \ln 60 = \frac{4.1}{18.7} = 0.22 \text{ ft/sec}$$

$$v = \text{gas velocity} = \frac{300,000}{60 \times 5 \times 27 \times \frac{8\frac{3}{4}}{12} \times 20} = 2.5 \text{ ft/sec}$$

17.5 APPLICATION TO OPEN HEARTH FURNACES

The open hearth operation is a batch steelmaking process carried out in a covered, refractory lined, rectangular furnace. The furnace is provided with a hearth in the shape of a shallow, rectangular dish into which hot metal from the blast furnace, scrap steel, iron ore, and limestone are charged. The charge is melted down and heated to temperatures as high as 3000°F by the combustion of such fuels as oil, natural gas, or tar with an excess amount of air preheated in regenerative heat exchangers called checkers. Two sets of checkers are normally associated with a furnace. Hot gas leaving the furnace is passed through one set of checkers, heating the refractory materials in the checker. The gas flow is periodically changed to the alternate set of checkers, and incoming air for combustion is heated by the other set. The hearth is called "open" because the charge is open to the sweep of hot combustion gases across its surface. The high temperatures, coupled with oxidizing ore and an excess of air, bring about the oxidation of the carbon and other elements to be removed from the iron. The process can be facilitated by oxygen lancing after meltdown, a practice which is followed in many open hearth furnaces. Limestone additions in the charge flux the impurities and hold them in the slag. A single batch or heat of steel may weigh as much as 500 tons and require from 7 to 12 hours to produce.

In practice, the open hearth is charged with cold scrap, iron ore, and limestone and the mass is heated by furnace burners. Molten pig iron is then added to the charge and the melting process continued until most of the cold material is molten. At this time, the oxygen lancing is started and continued until the desired carbon level is attained. Iron ore may be charged into the furnace during lancing to regulate the temperature. The burners are ordinarily shut off during oxygen blowing. When the carbon content is reduced to a sufficiently low value, the metal is alloyed by making appropriate additions to the furnace or ladle and the heat is complete.

Particulate emissions. The particulate matter carried out of the open hearth by the exhaust gases comes from a variety of sources, including dirt and other fines on the charge materials, oils, grease, and volatile metal oxides from the scrap charge. During oxygen lancing, large amounts of iron oxide are evolved and may be mixed with lesser amounts of nonmetallic oxides from the slag materials. Because the hearth reactions vary from one period to another, the exhaust gas, particulate load, and particulate composition likewise vary with the different periods of the heat.

During meltdown, the particulate from the open hearth consists mainly of dirt, oil, grease, and volatile metal oxides from the charge materials, plus any particulate that might be developed as a product of fuel combustion. The temperature of the gases leaving the furnace may be 3000°F or slightly higher. The temperature of the flue gases is reduced to 1200-1500°F in the checkers, and finally to about 550-650°F in the waste heat boilers before being sent to the air pollution control device.

During oxygen lancing, iron oxide is the principal particulate from the furnace. Normally, the furnace burners are turned off during this time and, although gas temperatures near the lance may be 3000-3500°F, the average temperature of the exhaust gas is 1200-1500°F. Again, these gases are reduced to 550-650°F in the waste heat boiler before being sent to the pollution control device.

A compilation of data on emissions from several open hearth furnaces, ranging in size from 60 to 600 tons, is given in Table 17.11.[4] Dust concentrations in the flue gases were usually measured after the gas left the checker work, and consequently do not reflect the presence of heavy particles deposited in the checkers. Data for this table were taken from several furnace operations using cold scrap and hot metal charges, with and without oxygen lances on the furnaces.

Table 17.11

Fume Emissions from Open Hearth Furnaces
(Adapted from Reference 4)

Furnace Type (Hot or Cold Metal)	Furnace Size (net) tons	Gas Volume scfm	Dust Loading gr/scf	
			Range	Avg.
hot	275	70,700 to 83,500	0.11 to 0.61	0.43
-	225	58,500	-	0.43
-	400	54,000	-	5.0
cold scrap hot pig	550-600	60,000 to 64,000 at 600°F	0.21 to 2.7	-
-	310	1,135,000 max. at 550°F	-	0.9
cold	60	14,400	0.1 to 2.0	0.69
cold	100	n.a.[a]	0.01 to 0.08	n.a.
cold	n.a.	n.a.	0.04 to 0.18	n.a.
hot	110	20,000	0.02 to 0.07	0.04
hot	n.a.	n.a.	0.1 to 1.4	0.43[b]
hot O_2 lanced	n.a.	n.a.	0.07 to 0.4	0.25[b]
hot	n.a.	n.a.	0.11 to 1.26	0.5[b]
hot O_2 lanced	205	40,000	0.1 to 1.2	0.6[b]
hot	225	33,000	0.5 to 2.5[c]	0.43[c]
hot	250	18,000 to 60,000	0.1 to 2.0	0.5
hot	250	n.a.	0.11 to 0.34	n.a.
hot	275	33,500 to 61,800	0.10 to 0.31	n.a.
hot O_2 lanced	330	37,700	0.8 to 2.5	n.a.

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[a] n.a. indicates that data were not available.

[b] Estimated by the reviewers.

[c] Range data are from an earlier reference than the average data; therefore, the average figure is probably more accurate.

The gas flow leaving the furnaces ranged from 14,400 scfm on the smallest furnace, to a maximum of 83,500 scfm on the larger furnaces during an active carbon boil. The dust load in grains per scfm of the flue gas varied not only between furnaces, but varied sometimes by a factor of 10 on the same furnace during different periods of the heat. The reported range of dust concentrations is from 0.01 to 2.5 gr/scfm.

An indication of the amount of variation in particulate emissions from the open hearth furnace during the different periods of the heat is given in Table 17.12.[4] The 60 net ton furnace, utilizing a cold charge and having an effluent gas volume of 14,400 scfm, had an emission rate of 0.87 gr/scf during charging. This decreased to 0.51 gr/scf during meltdown, and further decreased to a value of 0.34 gr/scf during working and refining. A larger 205-ton furnace with an effluent gas flow of 40,000 scfm maintained a particulate loading of 0.35 gr/scf during charging and meltdown. This increased to 0.45 gr/scf during the hot metal addition, and further increased to about 0.85 gr/scf during the carbon boil, and working and refining of the melt when an oxygen lance was used. Other furnaces using a combination of cold scrap and hot metal in the charge, and not employing an oxygen lance during the refining period, exhibited generally lower emission levels of 0.11 to 0.25 gr/scf, although emissions as high as 0.60 gr/scf were observed at times. Total emissions from the furnaces ranged from about 7.5 to 10 lb of dust per ton of steel when no oxygen lance was used, to a maximum of about 35 lb per ton with the use of an oxygen lance.[4,21,22]

The chemical composition of particulate from several open hearth furnaces is shown in Table 17.13. The particulate consists predominantly of iron oxide (Fe_2O_3) with minor amounts of silica, manganese oxide, aluminum oxide, zinc oxide, and lead oxide. The particulate contains over 60% iron, which is equivalent to a good grade of iron ore, but because of the zinc and lead contaminants, it is objectionable as a blast furnace charge material.

An indication of the variation in dust composition that may occur between the scrap charging and oxygen blowing periods of an open hearth heat is given in Table 17.14. During meltdown, the iron content of the particulate is lower, and zinc and lead contents are noticeably higher. In this case, a zinc content of 16.1% and lead content of 3.5%, are reported. During oxygen blowing, however, the iron content was near

Table 17.12

Variation of Fume Generation With the State of the Heat in Open Hearth Furnaces (Adapted from Reference 4)

Furnace Size (net) tons	60	--	--	205	--	--	--	275	550-600	--
Effluent Gas Volume, scfm	14,400	--	33,500 to 61,800	About 40,000	--	--	--	70,000 to 83,000	600,000 to 640,000	--
Stage of Heat				Dust Loading at 60°F and 29.9 in. Hg gr/ft^3						
Charging	0.87	0.11	0.27	0.35	0.17	0.24	0.1 - 0.58	0.56	0.78	0.1 - 0.58 $(.\overline{35})$
Meltdown	0.51	0.11	0.54	0.35	--	0.23	0.1 - 0.58	0.56	--	0.1 - 0.58 $(.\overline{35})$
Hot Metal Addition	only cold metal	0.33	0.39	0.45	only cold metal	0.25	0.24 - 0.66^2	0.61	1.9^2	0.24 - 0.66 $(.\overline{45})$
Ore and Lime Boil	--	0.43^1	0.60	0.82^2	0.18	0.25	0.51 - 1.20^2	--	2.7^2	0.14 - 1.6 $(.\overline{82})^2$
Working and Refining	0.34	0.66	0.19	0.87^2	0.18	0.40	--	0.11 - 0.18	0.21	0.51 - 1.2 $(.\overline{87})^2$

---

[1] Three values were obtained during lime boil; 0.43 is the median.

[2] High-purity oxygen was injected during this stage by roof lances.

Table 17.13

Open Hearth Dust Analyses
(from published data)

	Reference 23	Reference 25	Reference 26
Fe_2O_3	89.1 - 92.1		88.70
FeO	01.9 - 03.7		3.17
Fe	63.7	63.5 - 68	64.45
P		0.06 - 1.22	0.5
Mn		0.43 - 0.55	
MnO	0.50 - 0.60		0.61
SiO_2	0.30 - 0.90	1.16 - 1.56	0.92
Al_2O_3	0.10 - 0.50	0.15 - 0.44	0.67
CaO	0.70 - 0.80	0.68 - 1.06	1.06
MgO		0.32 - 0.44	0.39
Ca		0.11 - 0.16	0.14
Cr		0.06 - 0.11	
Ni		0.03 - 0.05	
Zn	0.40 - 1.70		
ZnO		0.26 - 2.04	0.72
Pb	0.10 - 0.50	0.05 - 0.95	
Cl		0.26 - 1.01	
Na		0.30 - 0.70	
S		0.33 - 0.70	0.92
C		0.06 - 0.12	
SnO_2		0.01 - 0.03	
Alkalies	0.20 - 1.40		

Table 17. 14

Analysis of Fume from Oxygen-Blown Steel Manufacturing Processes

	OH (A)		OH (B)	LD converter	
	Charging scrap	Oxygen blowing	Oxygen blowing	First five minutes	Average for blow
Iron dusts	44.0	78.4	89.6	83.1	83.1
Silica	0.4	0.5	0.5	1.5	1.2
Aluminum	6.8	7.95	6.1	1.0	1.3
Titanium	0.2	0.2	0.1	ND	ND
Manganese as Mn_3O_4	0.5	0.85	0.6	ND	ND
Calcium	1.0	0.60	0.6	2.9	3.3
Magnesium	0.4	0.17	0.0	ND	ND
Zinc	16.1	0.0	0.0	0.04	0.03
Lead	3.5	0.0	0.0	0.3	0.3
Phosphorus	0.7	0.6	0.7	ND	ND
Sulphate as SO_3	7.2	5.9	1.2	1.5	3.3
Dust concentration, gr/ft^3 at NTP	0.2-1.5 Av. 1.0	0.6-1.0 0.8	2.5-5.0 Av. 4.0	7.0-15.0 10.0	4.0-7.0 6.0
OH furnace (A)	Fuel, oil; charge 70% hot metal				
OH furnace (B)	Fuel, coke-oven gas; charge 100% hot metal				

80%, and the remainder was composed primarily of aluminum oxide. Zinc and lead contents of the fume were essentially zero during this period.

The size of particulate matter from open hearths ranges from less than 0.03 micron to several microns. One reported size distribution for a furnace without an oxygen lance is given in Table 17.15.[23] Figure 17.30 represents the particle size analysis of open hearth dust from a furnace utilizing an oxygen lance.[24] Composite samples representing dust evolved during the entire heat indicate that about 50% of the particulate is less than 5μ. During the lime boil, however, the fume is considerably finer, and as much as 77% of the particulate collected is less than 5μ in diameter and 20% less than 1μ. In number, a great majority of the particles are smaller than 0.1μ.[24]

Electrical resistivity of open hearth dust and fume is usually below the critical value of 2×10^{10} ohm-cm. In-situ resistivity data for several open hearth furnaces together with a typical laboratory resistivity-temperature curve are shown in Figure 17.31.

The composition of the waste gases from open hearth furnaces depends, to a considerable extent, on the type of fuel used to fire the furnace. Table 17.16 gives several stack gas compositions after the gas has been mixed with 40% excess air.[23] Producer gas, coke oven gas, fuel oil, coal tar, pitch, creosote, and natural gas, were used to fire the furnaces from which these waste gases were taken.

Three methods are in use for cleaning open hearth gases. These are: (1) the electrostatic precipitator, (2) bag house collectors, and (3) high energy wet scrubbers.

Open hearth precipitators. Figure 17.32 is a flow diagram for cleaning open hearth furnace gas with an electrostatic precipitator. Precipitators for open hearth gas cleaning are of the single stage, horizontal flow, duct type, with a steel shell. In modern plants, the open hearth shop is equipped with a multiplicity of furnaces, generally serviced by a common air pollution control system. In addition to the use of electrostatic precipitators for control, cyclone collectors and washers or scrubbers are used singly or in combination. Where a common system is used, the gas from all the furnaces is brought together in a collector main or mixing

Table 17.15

Size Distribution of Fumes in an Open Hearth Effluent
(Adapted from Reference 23)

Size-fraction (μ)	Percent in Size-fraction (weight)
1.0 - 3.0	7.3
0.5 - 1.0	28.4
0.15 - 0.5	49.5
Below 0.15	14.8

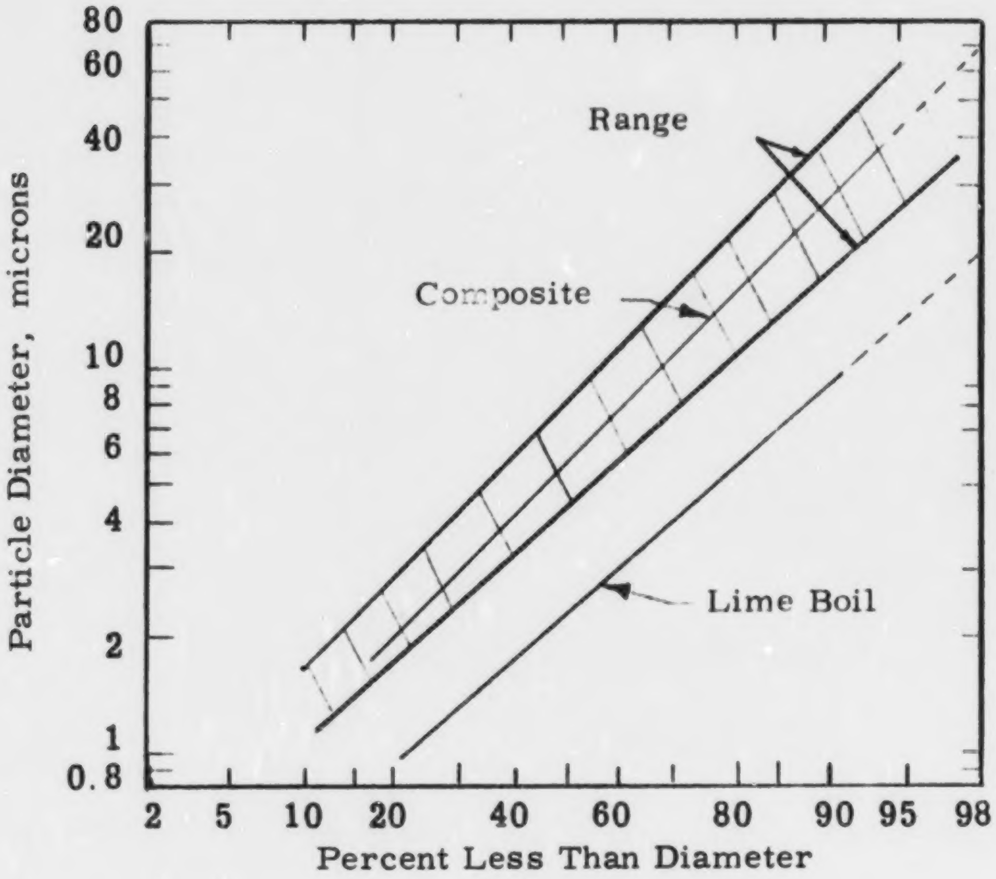

Figure 17.30. Particle Size Analysis of Open Hearth Dust from a Furnace Utilizing an Oxygen Lance.

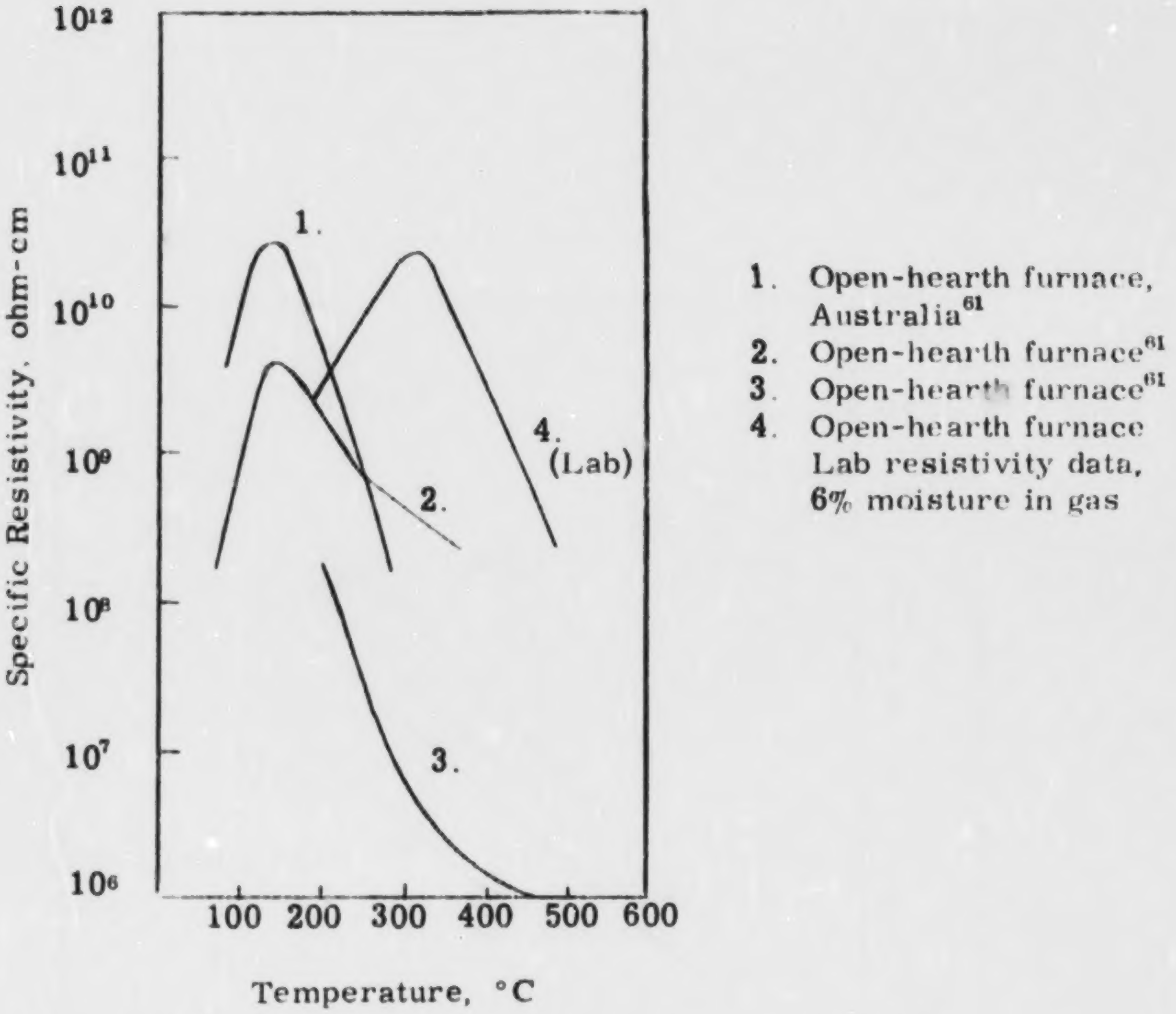

Figure 17.31. Electrical Resistivity of Red Oxide Fume from Three Oxygen Lanced Open Hearth Furnaces. A Laboratory Measurement at 6% Moisture is Also Shown.

Table 17.16

Waste Gases from Open Hearth Furnace Fuels
(Adapted from Reference 23)

	Flue Gas Composition (%) with 40% Excess Air					
Waste Constituent	Producer Gas	Coke Oven Gas	Fuel Oil	Coal Tar	Pitch Creosote	Natural Gas
CO_2	14.2	6.05	9.6	11.2	11.5	7.26
H_2O	6.9	16.4	13.4	11.3	8.9	13.45
N_2	74.8	72.0	71.5	72.0	73.95	73.75
O_2	4.1	5.4	5.4	5.4	5.6	5.54
SO_2	0.07	0.15	0.06	0.05	0.04	nil

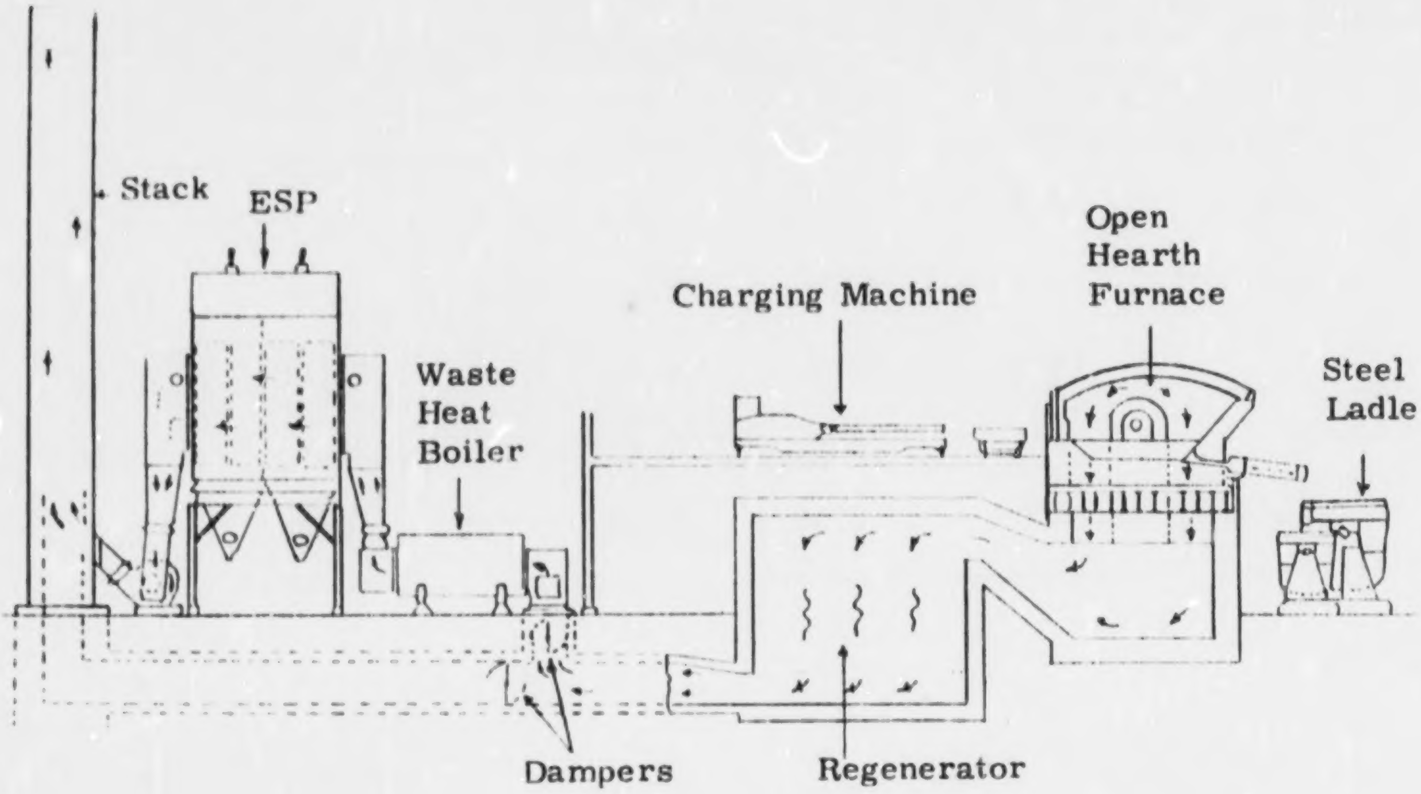
Stack
ESP
Waste
Heat
Boiler
Charging Machine
Open
Hearth
Furnace
Steel
Ladle
Dampers
Regenerator

Figure 17.32. Flow Diagram for Cleaning Open Hearth Furnace Gas With an Electrostatic Precipitator.

flue, in order to level out peaking of the individual furnace gas characteristics. Where needed, soot blowers are used in the bottom of the collector flue to keep the dust in suspension. The soot blower media are either compressed air or steam. Steam is used if electrostatic precipitator operation requires wetter gas in order to reduce the electrical resistivity of the dust. Open hearth shop practice employs a multiplicity of air pollution control systems spaced along the collector main. Each system is usually independently controlled, but overriden by collector main conditions.

In cases where air pollution control requires chemical treatment, such as for the reduction of fluorine emitted, additional equipment is needed. If a dry process is used, such as injecting lime into the gas, cyclone collectors are located ahead of the precipitator in order to give good mixing and longer retention time for the dry adsorbent. If a wet system is used, scrubbers or wash towers are located after the precipitator.

In operation, the gas temperature should be reduced to about 500°F for the electrostatic precipitator, to about 300°F for the high energy wet scrubber, and to the 250 to 450°F range for the bag house, depending on the kind of bags used. The electrostatic precipitators have the advantage of low pressure drop and an efficiency of 99.5% if properly designed and engineered.

Figure 17.33 is a schematic diagram showing a typical electrostatic precipitator installation on an open hearth furnace.

Precipitatator installations. Table 17.17 shows the number of electrostatic precipitators installed on open hearth furnaces during the period 1949-1968. Also shown in the table are the average design efficiencies for the precipitators installed during 5 year intervals. Figure 17.34 is a plot of the installed capacity of electrostatic precipitators for open hearth gas cleaning based on gas volume handled.

Precipitator design and performance. Design of precipitators for cleaning of open hearth effluent follows the general procedure outlined in Chapter 9, Part I. The size of the precipitator is governed by the collecting plate area, which is determined from the design efficiency, gas volume handled, and the precipitation rate parameter utilizing the Deutsch-Anderson equation. The precipitation rate parameter for open

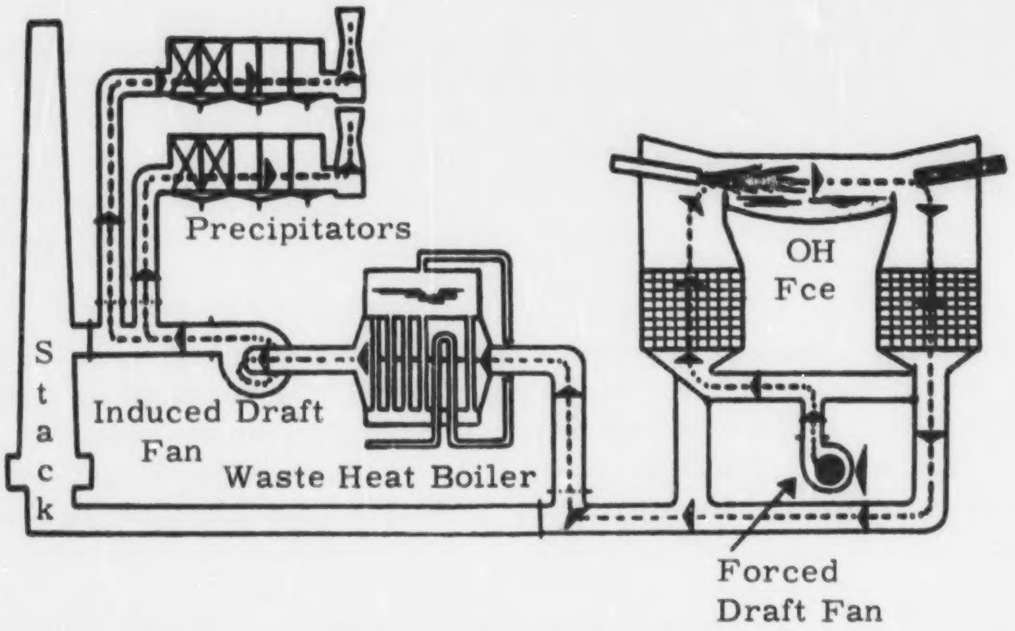
Precipitators
OH
Fce
S
t
a
c
k
Induced Draft
Fan
Waste Heat Boiler
Forced
Draft Fan

Figure 17.33 Schematic Diagram Showing a Typical Electrostatic Precipitator Installation on an Open Hearth Furnace.

Table 17.17

Precipitator Installations on Open Hearth Furnaces 1949-1968

Pptr. Contract Year	No. of* Install.	No. of Elec. Pptrs.	Total Gas Vol. (10^6 acfm)	Total Gas Vol. Accumul. (10^6 acfm)	Weighted Eff. on acfm Basis Yearly	Weighted Eff. on acfm Basis 5-Yr.	Five Year Period Total Gas Volume (10^6 acfm)	Five Year Period Average Vol/Yr during Period (10^6 acfm)
1949	1 (4)	4	0.134	0.134	95.5			
1953	1 (10)	8	0.400	0.534	96.0	95.9	0.534	0.107
1956	1 (11)	4	1.000	1.534	97.5			
1957	1 (1)	1	0.102	1.636	97.5			
1959	2 (12)	14	1.980	3.616	98.3	98.0	3.082	0.616
1960	3 (23)	13	3.229	6.845	98.7			
1961	1 (2)	2	0.330	7.175	99.0			
1962	2 (5)	4	0.382	7.557	98.4			
1963	2 (10)	3	0.453	8.010	98.1			
1964	1 (2)	1	0.260	8.270	99.2	98.7	4.654	0.931
1966	1 (7)	5	1.775	10.045	99.0			
1967	1 (8)	1	0.624	10.669	99.0			
1968	1 (9)	2	1.391	12.060	97.0	98.3	3.790	0.758
Grand Totals	18 (95)	62	12.060					

*Numbers in parenthesis are furnaces involved in installations indicated.

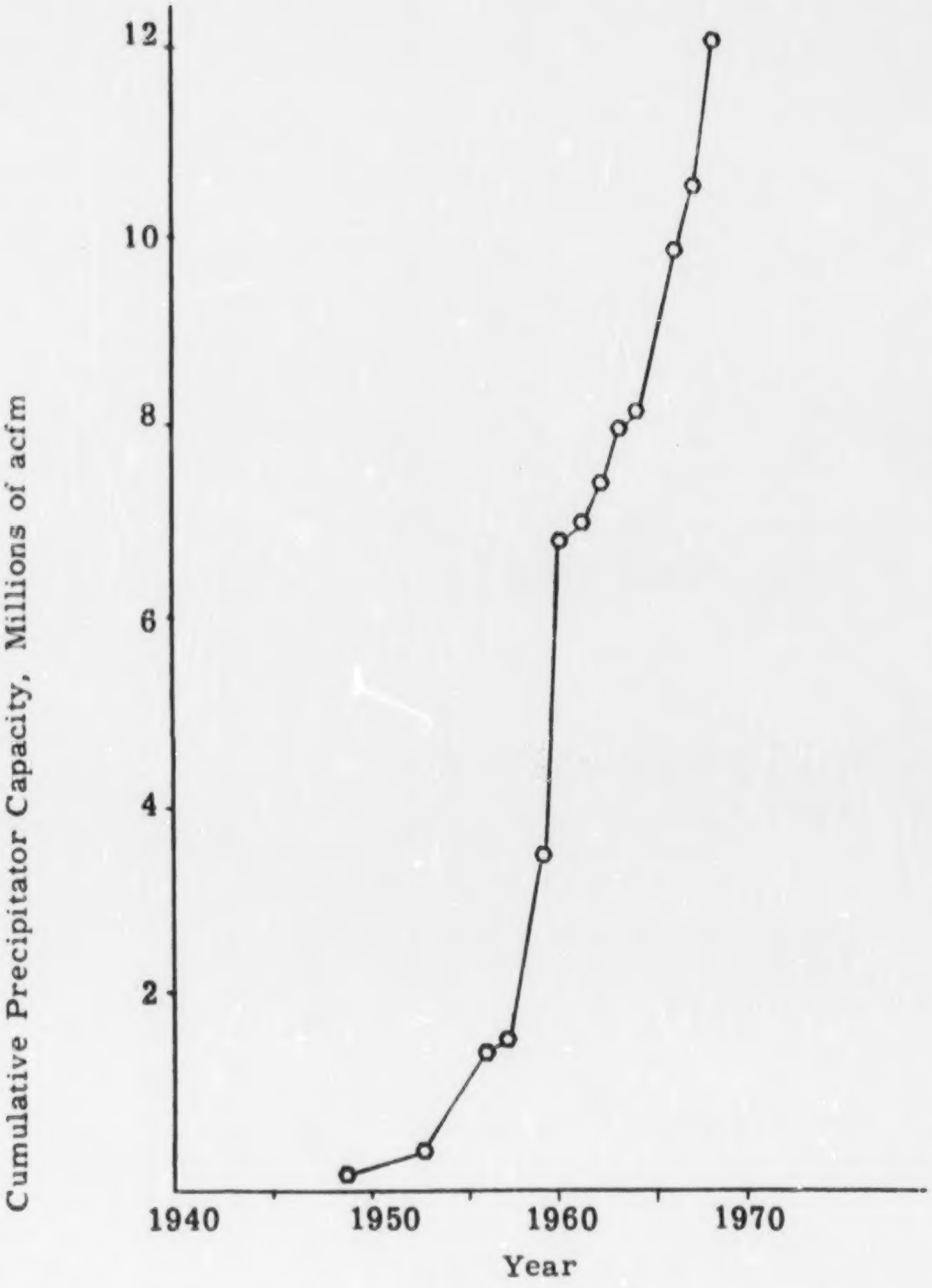

Figure **17.34**. Cumulative Installed Capacity of Electrostatic Precipitators for Open Hearth Installations.

hearth furnace precipitators varies with size and composition of the dust and effluent gas. Figure 17.35 shows the relationship between collection efficiency and the collecting area to gas volume ratio for a group of eight installations. The curves show a range of design w's of from 7 to 9 cm/sec. Test data are also plotted on the same curve.

Power supplied for a group of open hearth furnace precipitators is shown in Figure 17.36 in terms of installed power per 1000 cfm. Test data are also shown on the same figure. The trend is to supply more power than would normally be required. Also, power supplies are available in increments so that once power requirements are determined for a given application, the next modular size is chosen. The data, therefore, do not fall on a smooth curve.

Cost data. Capital and operating costs for open hearth precipitators are given in Table 17.18 for three different precipitator sizes.[14] Figure 17.37 shows the trend between cost and gas flow for a group of eight installations of varying size. The spread in the data is probably due to the variations in the particular installation more than any other single factor.

17.6 APPLICATION TO BASIC OXYGEN CONVERTERS

The basic oxygen process for producing steel utilizes high pressure oxygen introduced into the mouth of a basic, refractory lined converter to oxidize carbon, silicon, manganese, phosphorus, and other elements from the charge. Unlike electric furnaces or open hearths, there is no external source of heat for the converter, and consequently most of the charge must be composed of molten pig iron from blast furnaces. Oxygen is blown into the converter at high pressure through a water-cooled lance in volumes of up to about 25,000 scfm.[27] The high velocity oxygen stream impinges on the surface of the metal bath and produces violent agitation and intimate mixing of the oxygen with the molten charge. Rapid oxidation of dissolved carbon, silicon, manganese, chromium, and iron ensues. During the first part of the blow, the predominating reactions are between the oxygen and silicon, chromium, and manganeses. These metallic oxides enter the slag. After the slag-forming period, carbon oxidation increases until it reaches a peak at about the middle of the blow. If the phosphorous content is not too high, it may be removed during the normal blow. If the phosphorus is very high, however, as it is in some European irons, an after-blow or phosphorous blow may be required to remove this element from the charge.

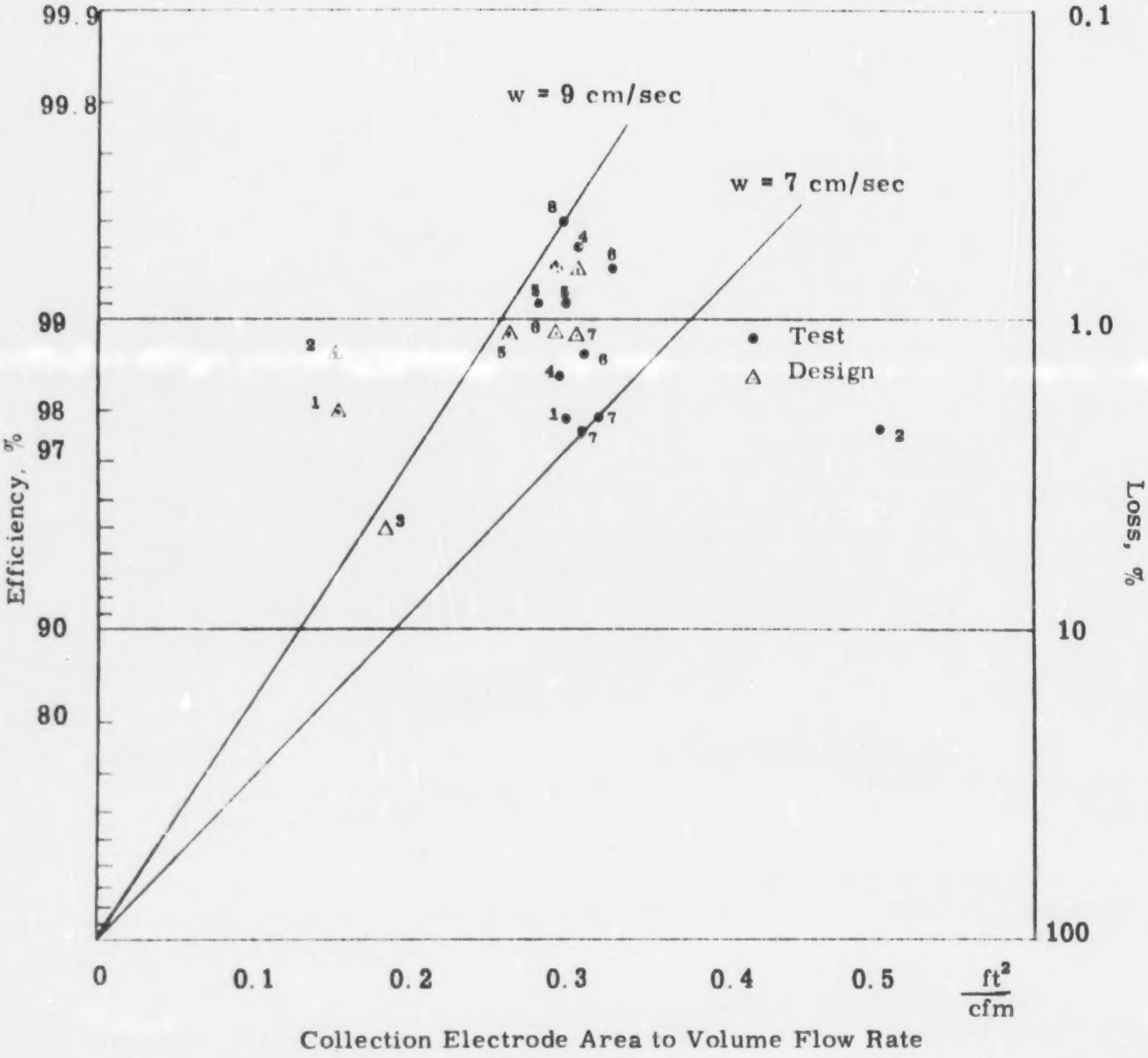

Figure 17.35. Relationship Between Collection Efficiency and Collection Electrode to Volume Flow Ratio.

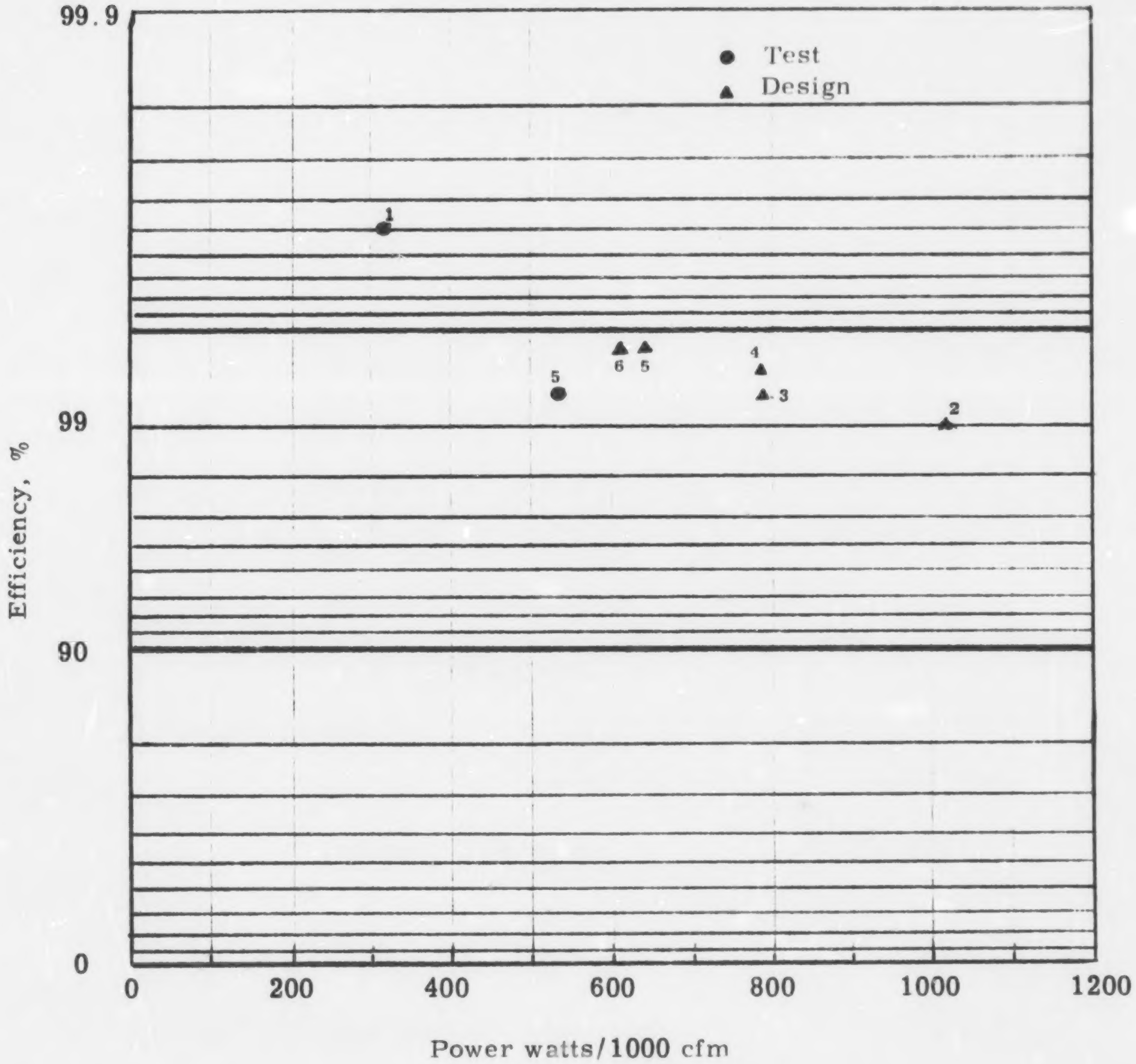

Figure 17.36. Relationship Between Installed Power Per Thousand cfm and Collection Efficiency for 5 Open Hearth Installations. Two Test Points also Included.

Table 17. 18

Open Hearth Furnace - Electrostatic Precipitator[14]

Gas Volume - acfm at 500°F	29, 000	85, 000	225, 000
	Capital Cost		
Furnace Size - tons	60	200	600
1. Material	$130, 000	$320, 000	$700, 000
2. Labor	70, 000	170, 000	380, 000
3. Central Engineering	52, 000	110, 000	200, 000
4. Client Engineering	13, 000	30, 000	50, 000
Total	$265, 000	$630, 000	$1, 330, 000
	Operating Cost		
1. Electric Power	$ 5, 000	$15, 000	$45, 000
2. Maintenance	11, 000	25, 000	54, 000
3. Operating Labor	20, 000	30, 000	40, 000
Direct Operating Cost	$36, 000	$70, 000	$139, 000
4. Depreciation	26, 500	63, 000	133, 000
5. Capital Charges	26, 500	63, 000	133, 000
Total	$89, 000	$196, 000	$405, 000

Note: 1) One Furnace System

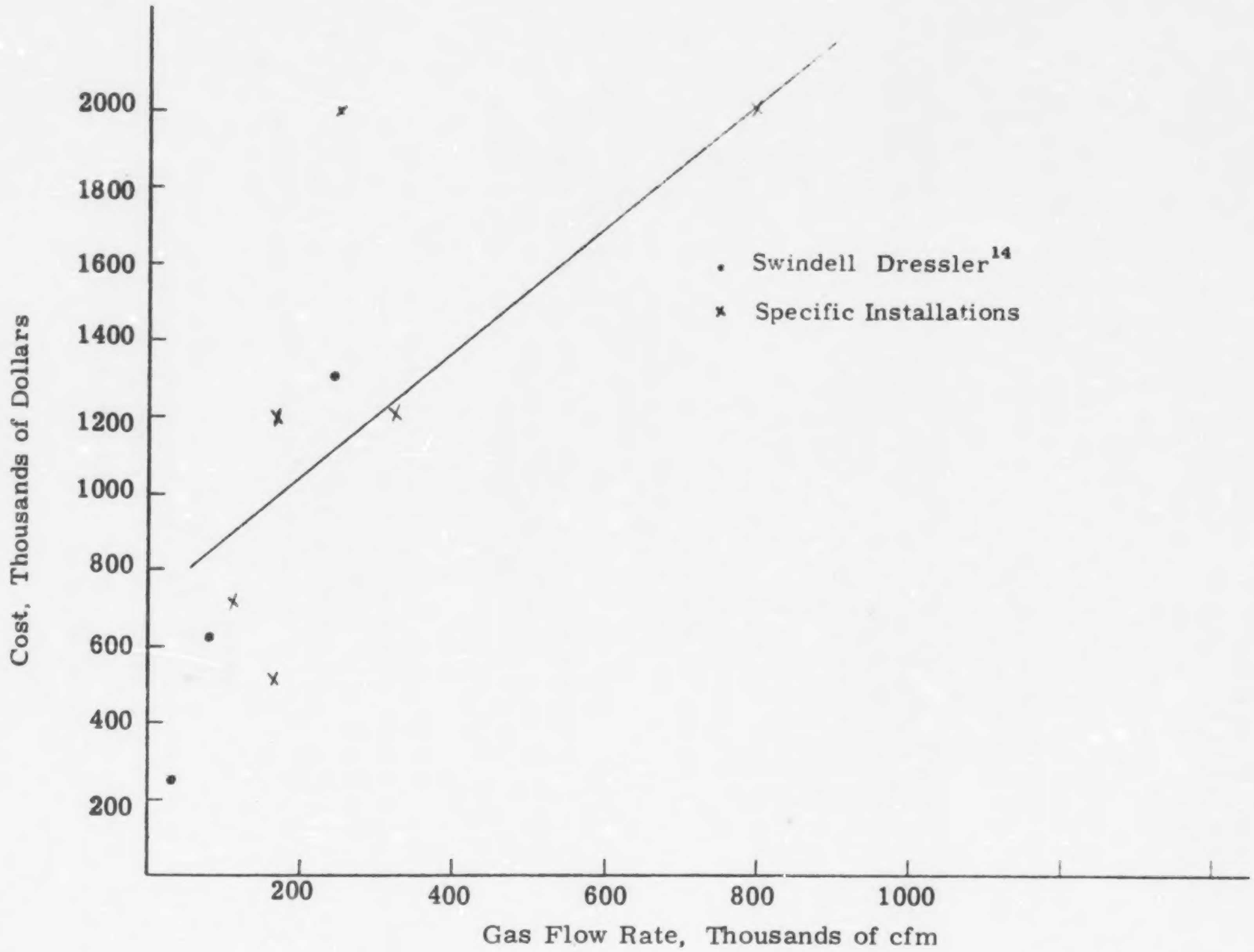

Figure 17.37. Relationship Between Costs and Gas Flow Rate for Electrostatic Precipitators for Eight Open Hearth Installations.

The furnace operation is cyclic and the time required for a complete heat averages about 45 minutes. The various steps in the cycle include charging, blowing, chemical analysis, alloying, and tapping. The charge usually contains about 75% molten pig iron, 25% steel scrap, and flux materials amounting to about 150 lb of lime, and 9 lb of fluorspar per ton of steel produced. The blowing time averages about 20 minutes, during which time large amounts of fume and particulate are evolved. The volume of gas generated may range from 200, 000 to 1, 200, 000 cfm at a temperature of 3000-3200°F, and may carry 300 lb or more of dust per minute from the mouth of the converter.[27,28]

Gaseous emissions from the BOF. Gas and fume emissions from basic oxygen converters vary greatly with the stage of the heat; thus the pollution control equipment must be designed for the peak periods rather than the average. Figure 17.38 presents test data obtained from BOF furnaces on gas emissions as a function of blowing time.[29] At a constant blowing rate, the gas evolved from the furnace during the first interval of the blow has a volume considerably lower than that expected from the assumption that the oxygen was completely converted to CO_2. However, with increasing time, the gas evolution increases and peaks out at a value substantially greater than the above expected value, assuming complete combustion. The reason for this behavior is that at the first of the blow, some of the oxygen enters the slag and metal to produce condensed metallic oxides. As the blow continues and the temperature of the bath increases, some of these oxides react with carbon in the melt to produce carbon monoxide gas, thereby producing the higher rate of gas evolution.[29] Dust control equipment must be designed for this peak rate of gas evolution.

The composition of the gas emitted from the BOF ranges between 70 and 90% carbon monoxide and 10-30% carbon dioxide through the blowing period.[29-32] However, at the peak of the blow, the carbon monoxide content may occasionally exceed 95%.[29] The temperature of the gas ranges from 3000 to 3200°F, depending on the scrap charge and cooling additives.[30]

The gas and fume from the furnace are collected by a closed hood over the mouth of the converter. There is usually a gap between the furnace mouth and the hood, which allows enough air to be drawn into the stream to completely burn the carbon monoxide to carbon dioxide. The gas is then cooled prior to cleaning.

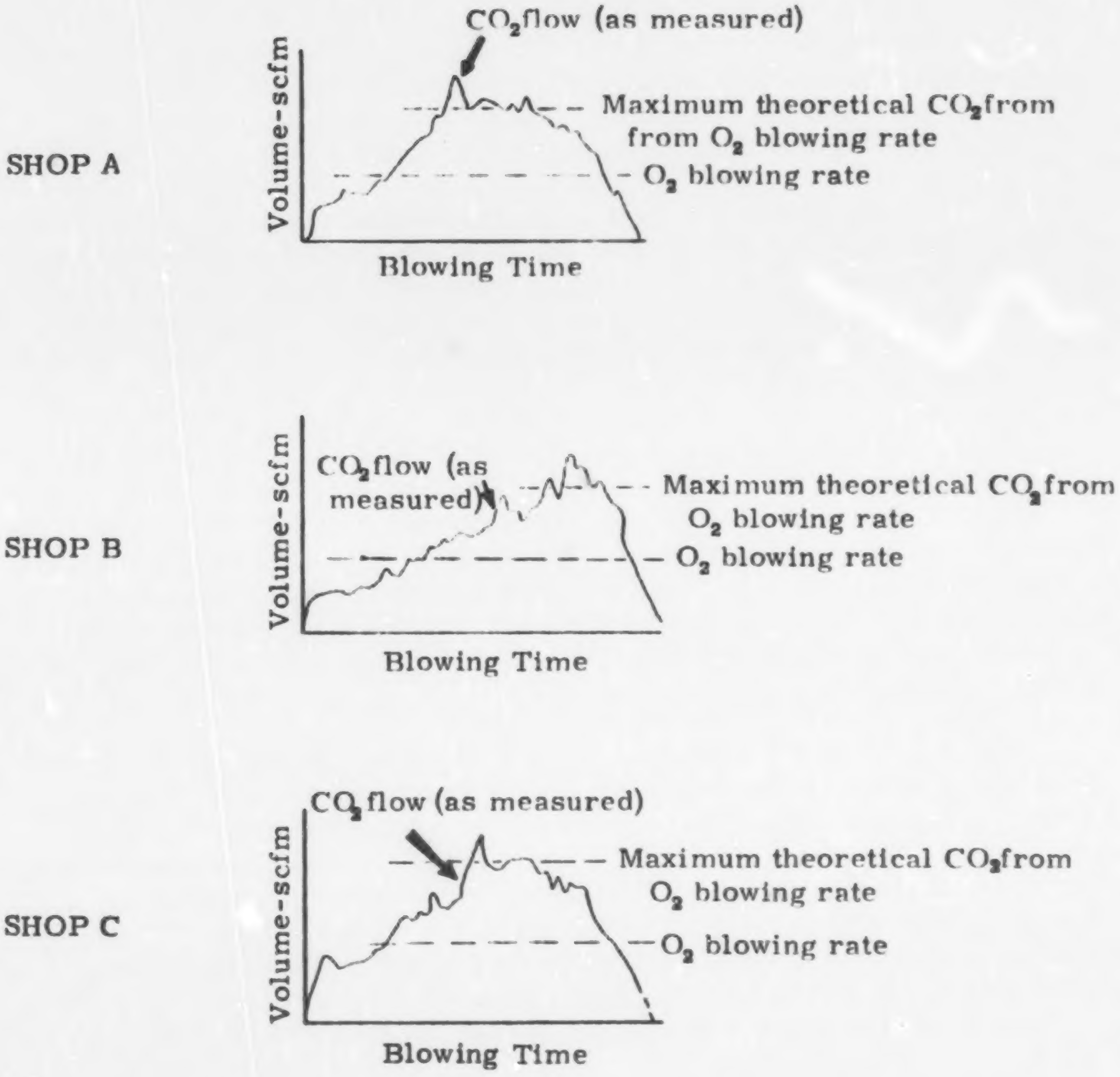

Figure 17.38. Carbon Dioxide from Basic Oxygen Furnaces After Complete Combustion.

Fume from the BOF. Fume from the basic oxygen furnace is primarily composed of the oxides of iron. The amount of fume varies greatly and reports show dust formation rates ranging from 20 to 60 lb/ton of steel.[32-34] This amounts to 1 to 3% of the metal produced and may produce concentrations of fume in the effluent gas of up to 15 grains per standard cubic feet under peak conditions,[35] with an average over the entire heat of about 5.5 grains per standard cubic feet. The dust is finely divided, with most of the particles ranging from 0.1 to 1μ in size,[27] although on a weight basis, a considerable portion may exist as relatively large particles. Two reported size distributions are given in Table 17.19.

The operation responsible for the production of such large amounts of fume is the high velocity oxygen lance, which produces violent agitation and rapid oxidation where it impinges on the molten bath. Temperatures of up to 3500°F, coupled with the absence of a slag cover and a boiling action of the metal which accompanies oxidation of carbon, cause the evolution of large amounts of fume.

The composition of two samples of dust from basic oxygen furnaces is shown in Table 17.20. Iron and its oxides comprise most of the dust. The analysis of one of the samples indicates that it is 90% by weight Fe_2O_3, while analysis of the second sample is reported to contain 66.7% metallic iron. The difference in degree of oxidation of the two dust samples is probably caused by the fact that the second sample was collected from the gas stream before the combustion of the CO, and hence the iron was not yet oxidized.

Resistivity of BOF dust is a function of moisture content in the temperature range in which surface conduction predominates. Figure 17.39 is a plot of resistivity as a function of gas temperature for BOF furnace dust. The resistivity values shown are laboratory measurements and cover a moisture range of from 5 to 25%. Since no moisture is released in the steelmaking process, all of the moisture comes from evaporative cooling of the gas in the cooling tower. Note that the resistivity is below the critical value of 2×10^{10} ohm-cm, indicating that no precipitator problems attributable to high resistivity are to be expected.

Gas cleaning of BOF effluent. Two cleaning devices are generally used for cleaning BOF effluent. These are high energy venturi scrubbers and electrostatic precipitators. As of 1968, there were 41 wet collectors

Table 17.19

Particle Size Analysis of BOF Fume from Two Installations

Reference 33		Reference 31	
Size - μ	%	Size - μ	%
<0.5	20	<5	8.9
0.5 - 1.0	65	5 - 10	9.1
1.0 - 15	15	10 - 20	39.9
		20 - 30	28.8
		>30	13.3

Table 17.20

Composition of Fume from BOF Data Shown in Table 17.19

Element	Reference 33 %	Reference 31 %
Fe_2O_3	90.0	4.01
Mn_3O_4	4.4	-
FeO	1.5	21.45
SiO_2	1.3	1.3
CaO	0.4	3.7
P_2O_5	0.3	0.36
Al_2O_3	0.2	-
Fe	-	66.69
C	-	-
MgO	-	tr
TiO_2	-	0.66
Cr_2O_3	-	0.11

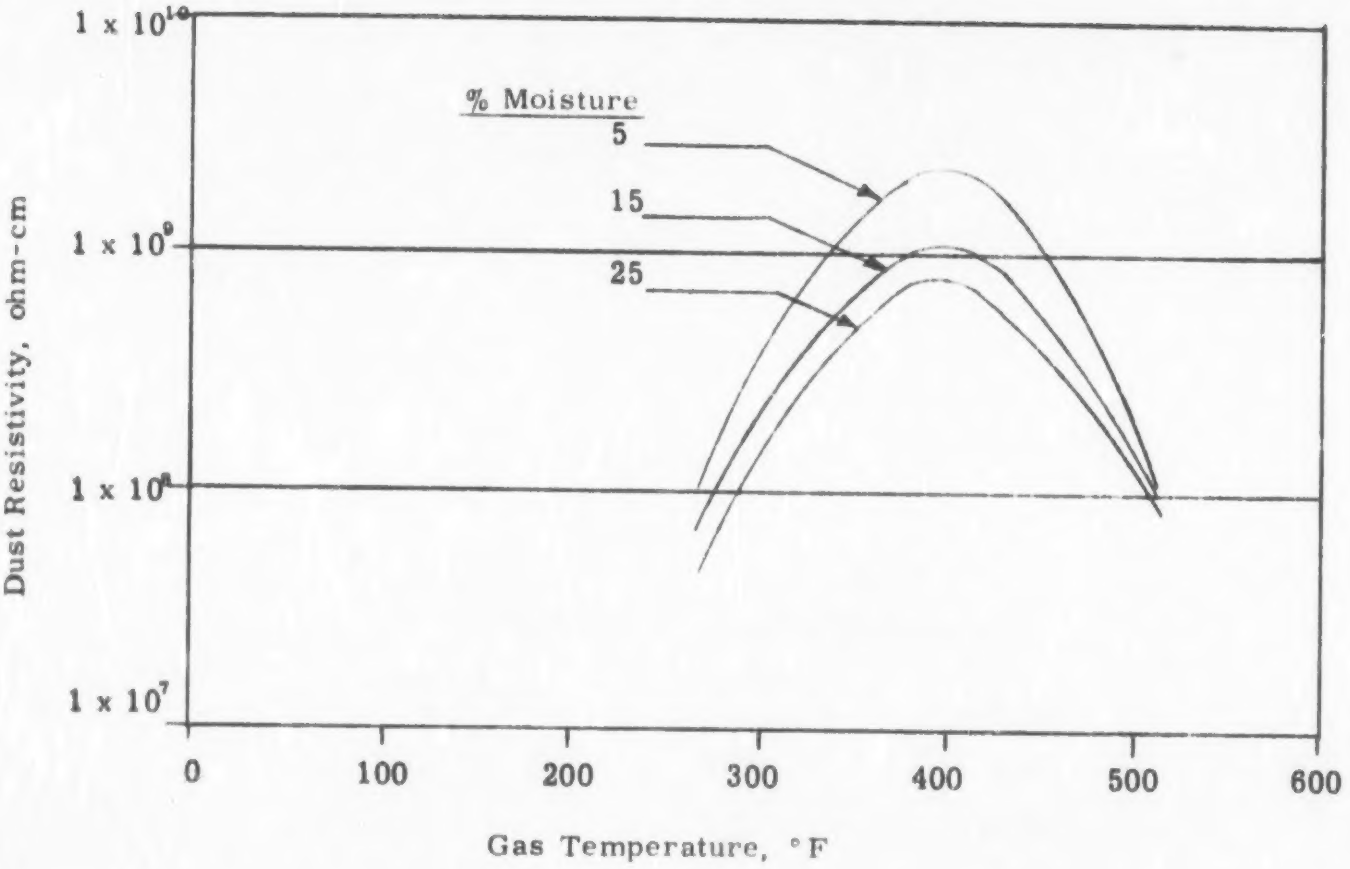

Figure 17.39. Resistivity vs. Gas Temperature for BOF Furnace Dust. (Laboratory Measurements)

and 36 electrostatic precipitator units on BOF furnaces in operation, under construction, or being designed.[29] A sketch of the electrostatic precipitator connected to the BOF is shown in Figure 17.40.

The disadvantages of the venturi scrubber are the high power requirements and the necessity for circulating and cleaning large amounts of water. The sludge water is passed from the venturi through a series of liquid cyclones which separate the coarser particles. Part of the overflow is returned to the venturi and the remainder flows to a clarifier-thickener, where particle settling may occur. The underflow is then pumped to a vacuum filter for dewatering. The difficulties with the water treatment plant include corrosion and buildup of suspended fume on moving parts. These problems may cause binding and wearing of the system because of the abrasiveness of iron oxide, which may lead to plant shutdown. Water scrubbing in a high energy system, however, may be very effective with collection efficiencies approaching about 99%.

The gas and fume must be cooled before being cleaned, regardless of the cleaning device employed. Cooling may be accomplished by passing the effluent over water-cooled heat exchangers or by passing it through a quencher where water sprays are employed. In the quencher, the coarse particles are washed from the gas stream and the gas is cooled to the temperature required before entering the cleaning device.

The water cooling tower provides the conditions for evaporation of water required to cool the gases, and hence moisture conditioning is a by-product of cooling and influences the resistivity of the dust in the temperature range in which surface conduction predominates.

Design of a cooling tower for cooling gases from metallurgical furnaces must be such that complete evaporation of the water takes place within the tower. If there should be water carryover to the electrostatic precipitator, the dust would tend to adhere to the plates, presenting rapping problems. Dust removal from the hoppers would also present a problem.

The factors that influence cooling tower design are the temperature of the gas, the size of the water droplets, and the residence time in the cooling chamber.

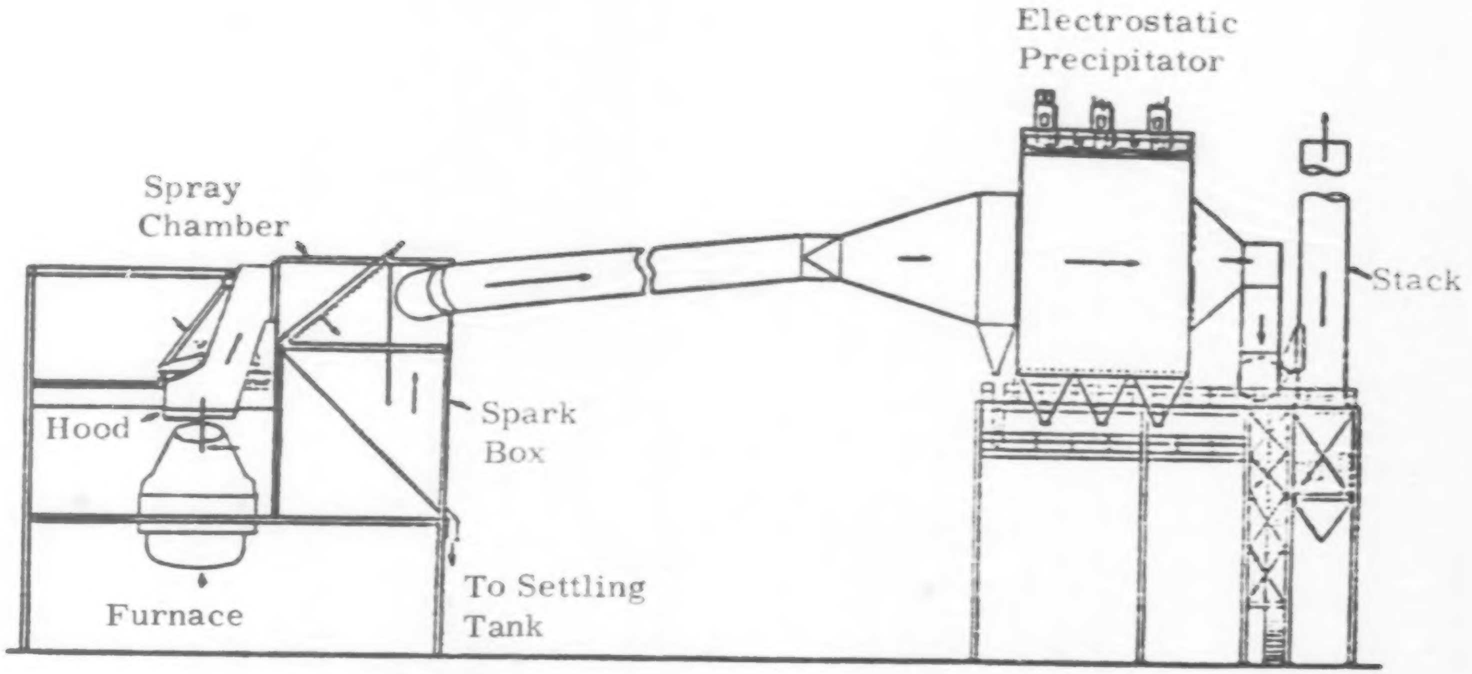
Electrostatic Precipitator
Spray Chamber
Stack
Hood
Spark Box
Furnace
To Settling Tank

Figure 17.40. Basic Oxygen Furnace Gas Cleaning With Electrostatic Precipitator.

Heat transfer rates are directly related to gas temperature so that a decrease in temperature would require greater surface area, greater residence time, or a combination of these to assure complete evaporation. Consequently, a cooling tower designed for exit gas temperatures of 500-600°F would be smaller than one designed for 200-300°F operation.

The surface area available for heat transfer is related to the size of the water drops. Since the gas flow in cooling towers is low, atomization takes place as a result of hydraulic pressure of the water. Higher water pressures generally result in smaller particle sizes. However, the effect decreases rapidly above a critical pressure level.

Residence time is a function of the volume of gas, the dimensions of the cooling chamber, and the gas flow distribution within the chamber. Whitehead and Darby[36] made a study of the effect of gas distribution in cooling chambers related to cooling incinerator gases. The same general conditions apply to the cooling of gases from metallurgical furnaces.

The dust loads and gas flows in BOF furnaces are extremely intermittent. During the oxygen blow, gas flow rates are high, whereas the dust level and gas flow are low between period of oxygen lancing. The cooling requirements, therefore, vary drastically during the steelmaking cycle. Because of the variations in temperature and gas volume, the amount of water required for gas cooling also varies over a wide range. There are two methods of controlling water flow rates. The water sprays can be arranged in banks and thermostatically controlled to come on at preset temperature levels. Alternatively, a spillback system can be used in which the rate of water injection is continuously varied over the full range of water flow rates.

If the cooling tower is designed with banks of spray nozzles for water rate control, the first bank of water sprays does not come on until a preset temperature is reached. Consequently, the moisture content of the gas is low and the dust resistivity is high during the interval between the start of the oxygen blow and initiation of water spray cooling. Unless provision is made to prevent it, abnormally high dust emissions can occur during this period. These emissions are called lance puffs.

Lance puffs can be controlled by the injection of steam into the duct ahead of the precipitator during the period before the water sprays come on. This practice is followed in many installations.

Precipitator installations. Table 17.21 shows the number of electrostatic precipitators installed during the period 1956-1968 and the gas volumes associated with the total installed capacity by year. Figure 17.41 is a plot of the total cumulative gas volume for the period 1956-1968.

Precipitator design. Precipitators for cleaning BOF dust are generally of the horizontal-flow, single-stage type. Mechanical components can be several types depending upon the particular application and can vary widely with the various precipitator manufacturers. Design efficiencies of precipitators for this service are high, consistently ranging above 99%.

The size precipitator required for a particular application is determined by the methods outlined in Chapter 9, Part I. The collecting plate area is calculated from the Deutsch-Anderson equation, which relates collection efficiency, gas volume, collection plate area, and the precipitation rate parameter.

Figure 17.42 is a plot of efficiency as a function of the collection plate area to gas volume ratio for a group of three BOF precipitators. The design values of precipitation rate parameters vary from 5.8 to 20 cm/sec for the particular units for which data were available.

Table 17.22 lists average values of some important design parameters for BOF furnace precipitators.

Cost data. Installed cost of BOF furnace precipitators range from \$4.88 to \$5.00 per cfm or \$8.00 to \$12.00 per sq ft of collecting surface. Figure 17.43 is a plot of erected costs for a limited number of installations.

Table 17.23 is a tabulation of estimated capital equipment costs prepared under the National Air Pollution Control Administration Contract PH-22-68-65 by Swindell-Dressler Company. The data for the three sizes of furnace are plotted in Figure 17.43 along with the Southern Research Institute data.

Table 17.24 gives the yearly operating costs estimated by Swindell-Dressler for those BOF precipitators handling three different gas volumes. The costs are broken down into categories of power, maintenance, operating labor, depreciation, and capital charges.

Table 17.21

Precipitator Installations on Basic Oxygen Furnaces 1956-1968

Pptr. Contract Year	No. of* Install.	No. of Elec. Pptrs.	Total Gas Vol. (10^6 acfm)	Total Gas Vol. Accumul. (10^6 acfm)	Weighted Eff. on acfm Basis Yearly	Weighted Eff. on acfm Basis 5-Yr.	Five Year Period Total Gas Volume (10^6 acfm)	Five Year Period Average Vol/Yr during Period (10^6 acfm)
1956	1 (2)	1	0.238	0.238	99.5			
1957	2 (3)	3	0.596	0.834	99.7			
1958	1 (1)	1	0.203	1.037	99.8	99.7	1.037	0.207
1960	1 (2)	1	0.242	1.279	99.5			
1962	2 (4)	6	1.630	2.909	99.2	99.2	1.872	0.374
1965	1 (1)	3	1.074	3.983	99.43			
1966	2 (3)	6	1.560	5.543	99.7			
1968	1 (2)	3	0.760	6.303	99.5	99.6	3.394	0.679
Grand Totals	11 (18)	24	6.303					

*Numbers in parenthesis are furnaces involved in installations indicated.

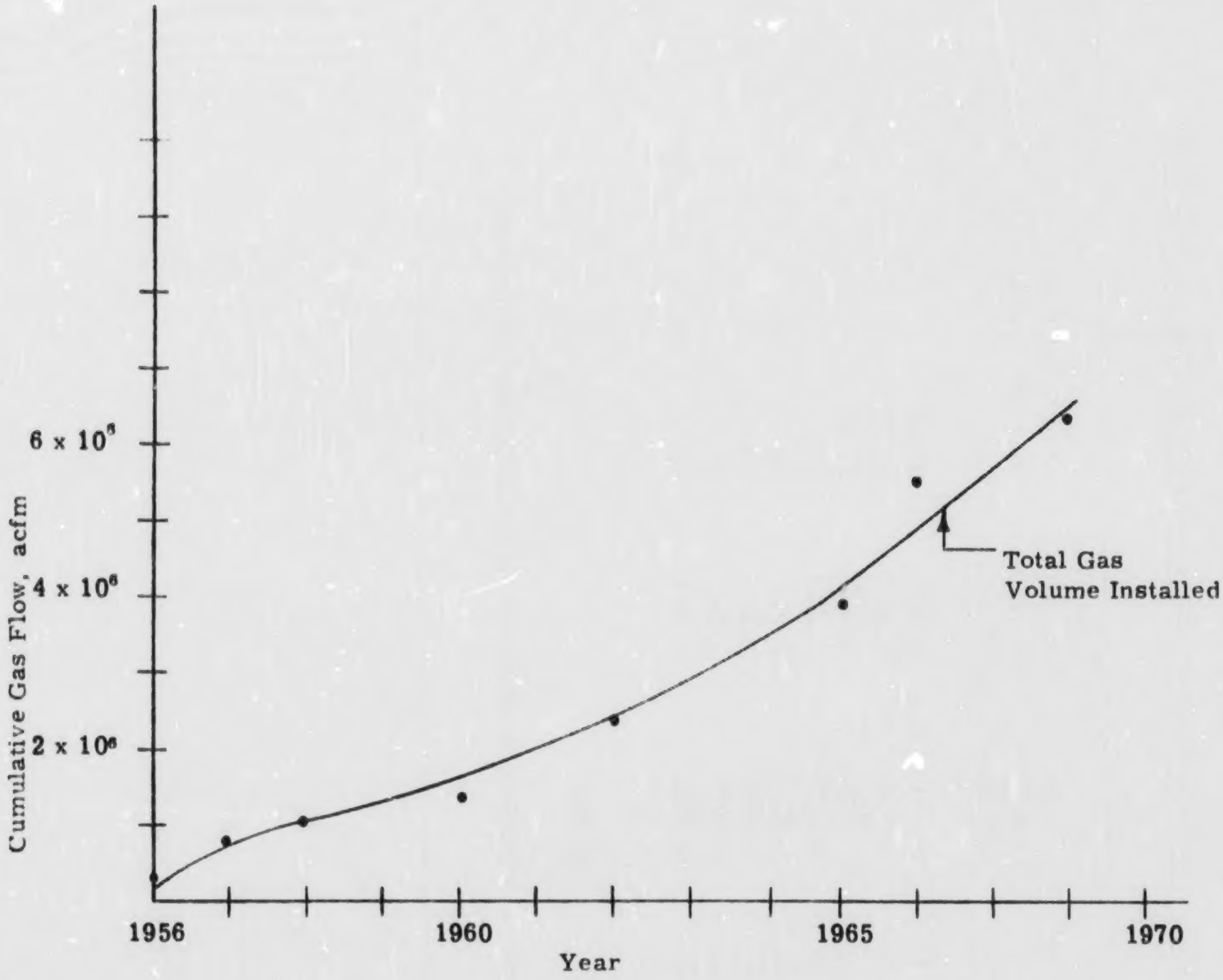

Figure 17.41. Cumulative Gas Volume from BOF Controlled by Electrostatic Precipitators.

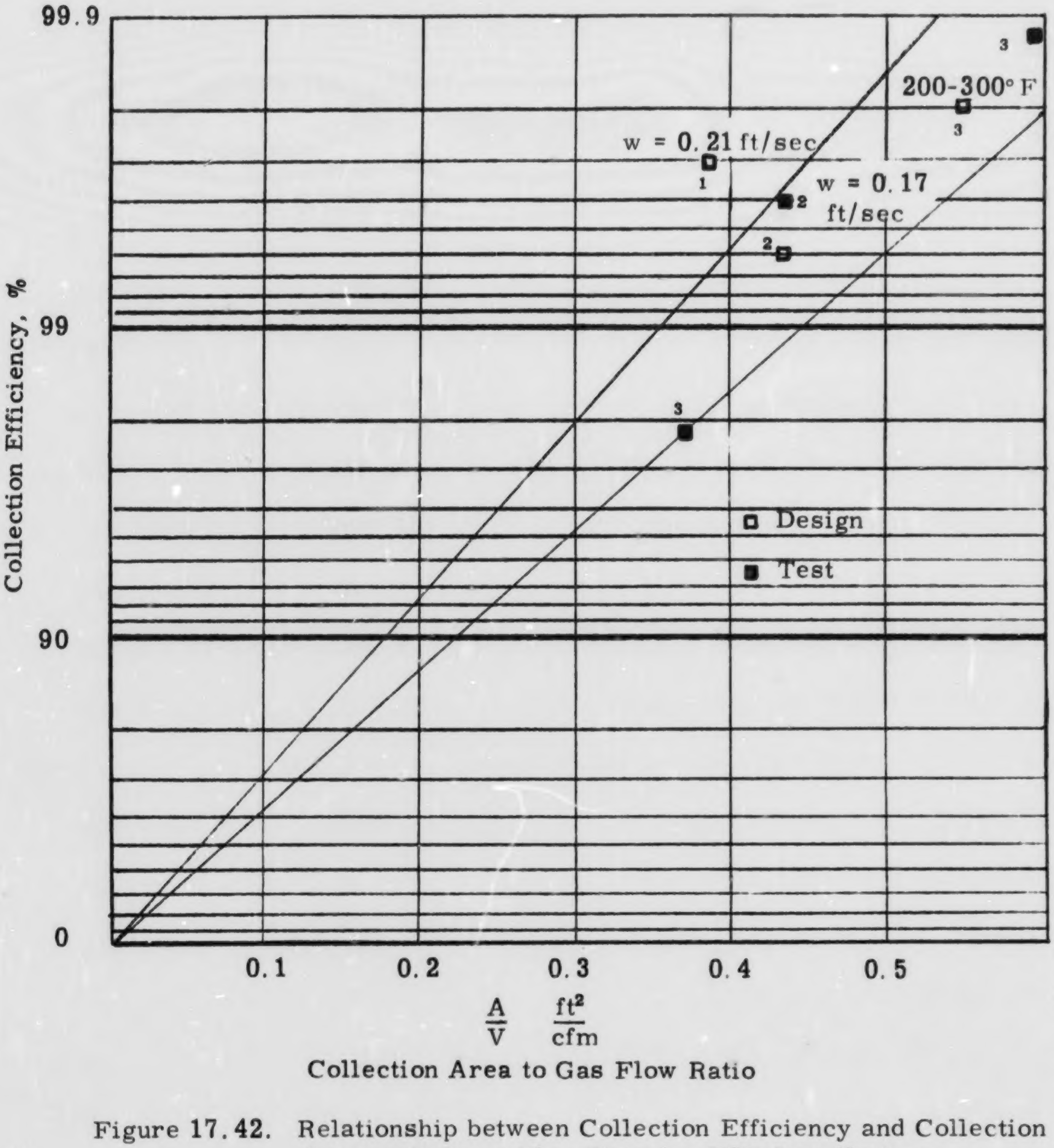

Figure 17.42. Relationship between Collection Efficiency and Collection Area to Volume Flow Rate for BOF Installation.

Table 17.22

Application and Design Variables for Two BOF Furnace Precipitators

Melt Rate	250 tons/heat	170 tons/hr
No. of Precipitators	3	3
Total Gas Volume, acfm	960,000	550,000
Gas Temperature, °F	550	230
Inlet Dust Loading, gr/cu ft	15	4
Design Efficiency, %	99.7	99.8
Collecting Surface Area/pptr.	146,200	111,500
Treatment Length, ft	42	36
No. Discharge Wires/pptr.	9,960	3,168
Length of Discharge Wires/pptr., ft	81,200	74,400
Diameter of Discharge Wires, in.	0.109	0.109
Rappers, Collecting Plates	32	36
Rappers, Discharge Electrodes	8	24
Power Input, watts/1000 cfm	304	215
No. of Sections	4	7
Design $\frac{A}{V}$, sec/ft	27.4	36.5
Design w, ft/sec	0.21	0.17

Table 17.23

Basic Oxygen Furnace - Electrostatic Precipitator
(Reference 14)

Capital Costs

Gas Volume - acfm at 500°F	375,000	785,000	1,200,000
Furnace Size - tons	100	200	300
1. Material	$ 900,000	$ 1,600,000	$ 2,250,000
2. Labor	450,000	800,000	1,100,000
3. Central Engineering	250,000	410,000	550,000
4. Client Engineering	60,000	100,000	140,000
Total	$1,660,000	$ 2,910,000	$ 4,040,000

Note: One Furnace System

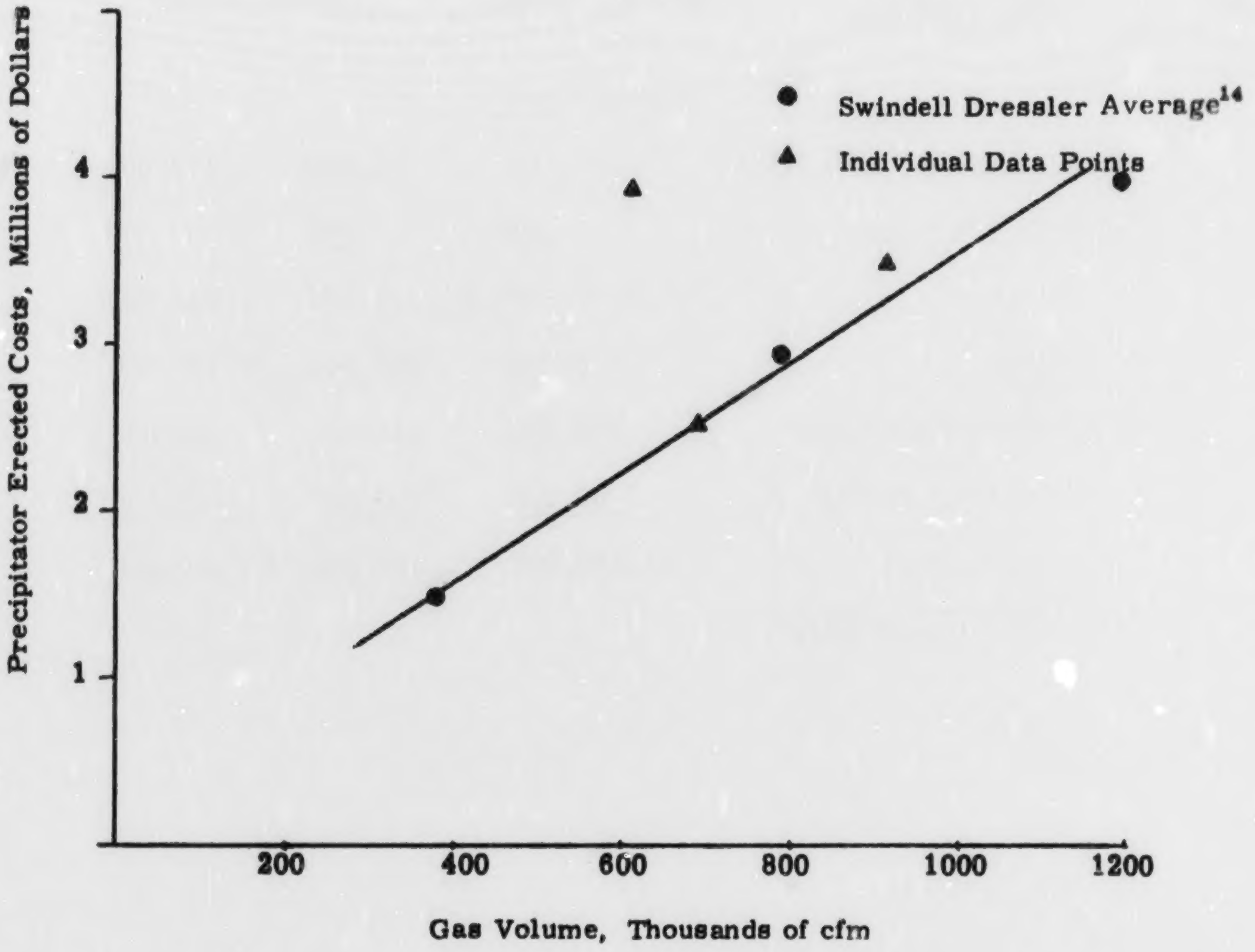

Figure 17.43. Erected Costs for a Limited Number of BOF Installations. Efficiency Range 99.4-99.8%.

Table 17.24

Basic Oxygen Furnaces - Electrostatic Precipitators
(Reference 14)

Operating Costs ($ /Yr)

Gas Volume - acfm at 500°F	375,000	785,000	1,200,000
Furnace Size - tons	100	200	300
1. Electric Power	$ 90,000	$ 210,000	$ 310,000
2. Maintenance	66,000	116,000	162,000
3. Operating Labor	20,000	30,000	40,000
Direct Operating Cost	$ 176,000	$ 356,000	$ 512,000
4. Depreciation	166,000	291,000	404,000
5. Capital Charges	166,000	291,000	404,000
Total	$ 508,000	$ 938,000	$ 1,320,000

Note: One Furnace System

17.7 APPLICATION TO ELECTRIC ARC FURNACES

Electric arc furnaces employed for melting and refining steel are cylindrical, refractory-lined structures, with a dish bottom and a domed roof. They range in size from a diameter of about 7 feet, with a hot metal capacity of about 4 tons, to a diameter of 24.5 feet with a metal capacity or 200 tons. The refractory lining used inside the steel structure may be either acid or basic, but basic refractories are more commonly used. Because of the flexibility of operation, the electric arc furnace is used to produce steels with a wide range of composition, including carbon, alloy, and stainless grades. In 1967, 12% of the steel produced in the United States was produced in electric arc furnaces.[37] A flow diagram of the electric arc process and auxiliary equipment is shown in Figure 17.44.

In practice, steel scrap, and perhaps hot metal from a blast furnace, are charged into the furnace, the electrodes lowered, and the power turned on. Power for melting and super-heating the charged material is supplied to the furnace through externally supported carbon electrodes which are automatically raised or lowered through holes in the furnace roof. Intense heat is produced by the current arcing between the electrodes and the metal charge, and this, coupled with the resistance heating that occurs as the current flows through the charge materials, results in melting and super-heating of the charge.

Meltdown in usually accomplished under oxidizing conditions in order to effect removal of phosphorus, and achieve a good carbon boil after complete meltdown. Oxygen for the oxidizing reactions, which occur before and during the carbon boil, may be obtained from the following sources:

1. oxygen gas injected into the bath after meltdown (oxygen lance),
2. oxygen from the furnace atmosphere,
3. decomposition of limestone, if this is used in the charge.
4. oxides of alloying elements used in the charge, and
5. iron ore or mill scale used in the charge or added after meltdown.

The use of an oxygen lance is important in modern steelmaking practice because of the rapidity with which the carbon boil may be initiated, and oxidizable elements removed from the melt.

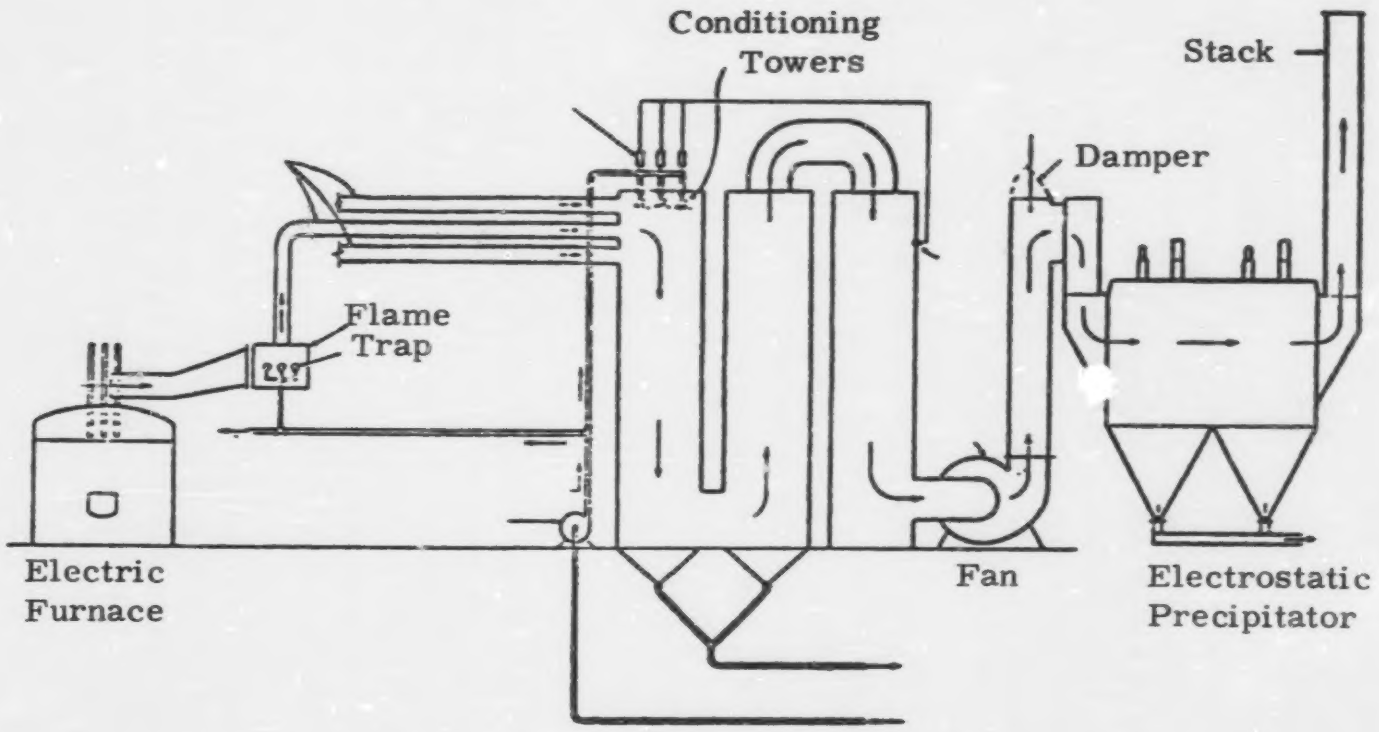
Conditioning
Towers
Stack
Damper
Flame
Trap
Electric
Furnace
Fan
Electrostatic
Precipitator

Figure 17. 44. Flow Diagram for Cleaning of Electric Furnace Exhaust Gases.

In production of low carbon steels, the temperature of the bath is increased during the oxygen lancing in order to promote carbon removal, and to increase the fluidity of the bath. The reaction of oxygen with carbon in the melt forms CO gas, and this gas gives rise to the carbon boil which is beneficial in purging the bath of hydrogen and nitrogen. Maximum cleanliness of the steel requires a hot bath with a vigorous carbon boil.

In the double slag refining practice, the oxidizing slag is removed from the furnace after the carbon boil is complete, and a basic deoxidizing slag is then prepared. The second slag facilitates sulfur removal from the molten metal, while the temperature and composition of the metal are being adjusted before tapping. As rapidly as possible, the composition and the temperature of the metal are adjusted to the desired range. After the proper temperature and composition are obtained, the power is turned off, the electrodes raised, and the furnace is tilted to allow the steel product to pour from the furnace into a collection vessel.

Gas emissions. Very little air is admitted to the electric arc furnace during the meltdown period, and during this time the furnace operates with an atmosphere containing a small amount of carbon dioxide, some hydrogen, but predominately carbon monoxide.[38] The particulate emissions from the electric arc furnace during this meltdown period are composed primarily of volatile matter from the charged scrap including oil, grease, and oxides of metals with high vapor pressures. Zinc oxide from galvanized scrap is probably the most common metallic oxide evolved during this period.

After meltdown of the complete charge, oxygen is lanced into the furnace. This oxygen combines with carbon, silicon, and manganese in the melt, to produce a furnace atmosphere containing primarily carbon monoxide, with some free oxygen, carbon dioxide, and hydrogen. Tests have indicated that the atmosphere in the furnace during lancing contains in excess of 80% carbon monoxide.[39]

The approximate temperature of the dust-laden gas leaving the electric arc furnace has been reported to be in the 2700°F to 2900°F range,[40,41] although some investigators have reported lower exit temperatures of about 1800°F during melting, and up to 2300°F during oxygen lancing.

The discrepancies in reported temperature of exit gases are probably caused by varying amounts of ambient air being entrained in the gas, either before or after the gas exits from the furnace shell.

The flue gases contain large carbon monoxide concentrations when released from the electric furnace. Because this constitutes a potential explosion hazard, the gases must be burned or significantly diluted with air before they enter a precipitator or other dust cleaning device. This is usually accomplished by admitting air into the stream of hot gases and allowing combustion to take place.

The flue gas volume to be treated with a dust collecting device ranges from a minimum of 12 to perhaps 200 times the volume of oxygen lanced into the furnace, depending on the method used to cool and collect the waste gases. A minimum value of 12 to 15 times the volume of oxygen lanced is obtained when using a direct furnace shell extraction followed by combustion, and subsequently, a water spray cooling to saturation at about 175°F.[40] In operations where a shop roof hood is used to collect fumes, it may be necessary to handle gas volumes as great as 200 times the volume of oxygen lanced into the furnace in collecting devices. Because of the wide variety of methods used in collecting fume, the actual dust load, gas composition, and gas temperature entering a dust collecting device varies greatly from plant to plant.

Particulate emissions. A compilation of data on total particulate emissions from various electric arc furnaces is given in Table 17.25.[42] The data presented are from both acid and basic electric arc furnaces with various melting rates, cycle times, and with furnace sizes ranging from 3 to 50 tons. The average quantity of fume is 10.6 pounds per ton of steel, or about 0.5% of the metal charged into the furnace. The range is from 4.5 to 29.4 pounds of particulate per ton of steel melted. However, emissions as high as 37.8 pounds per ton have been reported,[42] and occasionally, with highly oxidized scrap charge materials, emissions as high as 75 pounds per ton are reported.[38]

The wide variations in total fume emissions per ton of metal melted are attributable to several factors including: the size of the furnace and the melt practice employed, formulation of the charge, cleanliness of the scrap, melt refining procedure, and pouring temperature. The most

Table 17.25

Electric Arc Steel Furnace Fume Emission Data

Case	Rated Furnace Size, tons	Average Melting Rate, tons/hr	Cycle Time, hr	Fume emission/ton Melted, lb/ton	Furnace Process	Reference
A	50	18.3	4	9.3[3]		
	75	23.5	4	18.6[4]	Basic, single slag	44
	75	23.5	4			
B	50[1]	14.4	4	7.6		
	50[1]	13.6	4	6.9	Basic, single slag	42
	75[1]	21.9	4	12.3		
	3	1.5	2	12.6	Acid, oxygen blow	
	3	1.1	2	7.6	Acid, oxygen blow	
	6	3.1		10.4		
	10	6.6	2	5.5	Basic, oxygen blow	
	10	5.4	2	5.2	Basic, oxygen blow	
	2[2]	1.5[2]		13.4		
C	3	1.9	2	4.5	Acid	43
	3	1.6	2	5.8	Acid	
	3	1.9	2	5.7	Acid	
	6	2.6	2.3	15.3		
	6	3.0	2	12.8		
D	18	5.4	3	6.1		
E	6	4.1	1.2	29.4	Acid, single slag	42
F	3	1.8	1.8	12.7	Acid, single slag	51

---

[1] Refer to same furnaces as Case A.

[2] Two 2-ton furnaces operating in parallel.

[3] Average for one 50-ton and two 75-ton furnaces processing normal scrap.

[4] Average for one 50-ton and two 75-ton furnaces processing dirty, subquality scrap.

variables affecting effluent are probably quality and cleanliness of scrap, and the metal-refining procedure.

The types of scrap charged is quite important in determining the total emissions from the furnace, since the inclusion of large quantities of low boiling point, nonferrous metallic impurities in the charge will inevitably lead to high concentrations of oxides of these metals in the fume.[43] The cleanliness of the scrap is also significant because volatile impurities, including oil and grease, will be driven from the charge materials at relatively low temperatures during the early stages of meltdown, producing significant amounts of carbonaceous particulate matter.[44]

Quantitative tests designed to determine the effect of scrap cleanliness on the total particulate effluent have been performed using identical furnace operating conditions, with variations occurring only in cleanliness and quality of the scrap. The amount of fume emitted per ton of metal increased by a factor of 2 when dirty, subquality scrap was used.[44] The work of Kane and Sloan[43] substantiates these results, and reports an increase of over 40% in the amount of fume released per ton of metal processed when lower quality scrap was charged into the furnace.[43]

After meltdown and evolution of the readily volatile scrap constituents, the rate of fume release increases and reaches a peak during the boil and refining periods of the heat, particularly when an oxygen lance is employed.[45] If a mill-scale boil is employed, rather than an oxygen lance, the amount of fume produced may be somewhat less because of the less vigorous boiling action in the furnace,[45] and the lower melt temperature.[46] Also, an appreciable amount of fume produced during a mill-scale-induced carbon boil may be entrapped in the slag cover, whereas fume from an oxygen lanced heat may escape the slag cover because of the greater turbulence. During the refining phase of producing the steel, the fume rates are generally high, and tend to peak toward the end of the carbon boil. Metal oxide fumes from the melt normally decrease after the slag is shaped up, because the slag blanket tends to entrap them.[44,47]

Just as the amount of particulate matter from the electric furnace varies with the charge and operating conditions, so does the composition of the particulate. Table 17.26 presents some average chemical analyses of particulate emissions obtained from five alloy and stainless steel heats

Table 17.26

Chemical Analysis of Fume Samples

(Filtered from five alloy and stainless heats during shell evacuation test by Firth Sterling, Inc.)

Analysis for	Percent by weight
Fe_2O_3	23.95
FeO	9.66
CaO	15.41
MgO	15.25
Soluble Alkalis	11.25
Cr_2O_3	8.77
SiO_2	3.76
MnO	2.75
C	1.60
MoO_3	0.95
S	0.64
NiO	0.62
V_2O_5	0.51
Al_2O_3	0.47
CO_3O_4	0.08
TiO_2	0.05
N_2	0.014
P_2O_5	Tr.
Moisture	0.90
Ignition loss (probably carbonates)	0.35

produced in an arc furnace.[41] These data indicate that the fume is predominantly composed of the oxides of iron, calcium, magnesium, chromium, silicon, and the alkali elements. The range in composition of dust samples from an electric furnace shop is given in Table 17.27, and probably represents the maximum variations that might occur.[48]

Some qualitative spectrochemical analyses and concentration estimates of minor constituents in fumes from electric arc furnaces are given in Table 17.28.[41] The oxides of zinc, manganese, and lead are the predominant minor constituents with zinc probably coming from galvanized scrap, lead from terne sheet or free machining steels, and manganese from almost any type of scrap. Significantly higher zinc oxide contents have been reported, however,[11,42] and one such analysis is given in Table 17.29. The presence of 37% zinc oxide in this particulate sample is perhaps the result of a large amount of galvanized scrap being included in the charge.

The composition of fume changes during the different periods of melt production. Fumes may contain 40 to 50% iron by weight during meltdown, particularly if thin sheet scrap is used as a charge material, and may maintain this value during the carbon boil.[47,49] The remainder of the particulate evolved after meltdown is composed of the oxides of calcium, silicon, phosphorus, manganese, and sulfur.[50] After the slag is shaped up on the metal bath, the iron oxide content may fall to as low as 5%, with the calcium oxide content of the fume increasing to the 45 to 50% range.[49]

A representation of the dust burden during oxygen lancing of an electric arc furnace with a 15-ton capacity is given in Figure 17.45. During the meltdown period, a dust burden of up to about 3 grains per cubic foot was observed after the gas had been extracted from the arc furnace, combusted to remove the high carbon monoxide concentration, and water spray cooled to 175°F. During oxygen lancing, the particulate concentration was approximately 3 grains per cubic foot until 70% of the oxygen had been blown. During the last stages of blowing, the particulate burden decreased to about 1 grain per cubic foot.

Dust from electric arc furnaces tends to be extremely fine as indicated in Table 17.30.[41] These data, from several sources, indicate that 90 to 95% of the fume is below 0.5 micron in size. This very fine fume restricts the method of collection to electrostatic precipitators, bag-houses or very high energy wet scrubbers. The data presented in Table 17.30 suggest that there is a wide variation in particle size for

Table 17.27

Chemical Analysis Range of Electric Furnace Dust

Component	Range, %
Fe_2O_3	19-44
FeO	4-10
Total Fe	16-36
SiO_2	2- 9
Al_2O_3	1-13
CaO	5-22
MgO	2-15
MnO	3-12
Cr_2O_3	0-12
CuO	<1
NiO	0- 3
PbO	0- 4
ZnO	0-44
Alkalis	1-11
P	<1
S	<1
C	2- 4

Table 17.28

Qualitative Spectrochemical Analysis and Concentration Estimates of Detected Constituents
(Based on three typical samples of fume removed from hoppers of cloth filters.)

	Plant A	Plant B	Plant C
Element	Concentration Range, %		
Iron	5 to 50	Major Const.	Major Const.
Calcium	3 to 30	0.05 to 0.50	0.05 to 0.50
Silicon	0.10 to 1.00	2.0 to 20	2.0 to 20
Zinc	0.01 to 0.10	0.30 to 3.0	0.05 to 0.50
Manganese	0.50 to 5	0.20 to 2.0	0.20 to 2.0
Magnesium	0.50 to 5	0.010 to 0.10	0.005 to 0.05
Sodium	0.50 to 5		0.010 to 0 10
Chromium	0.30 to 3.0	0.5 to 0.50	0.10 to 1.00
Nickel	0.05 to 0.50	0.02 to 0.20	0.05 to 0.50
Aluminum	0.05 to 0.50	0.20 to 2.0	0.05 to 0.50
Lead	0.05 to 0.50	0.20 to 2.0	0.01 to 0.10
Molybdenum	0.05 to 0.50	0.005 to 0.05	0.005 to 0.05
Copper	0.02 to 0.20	0.03 to 0.30	0.03 to 0.30
Tin	0.01 to 0.10	0.01 to 0.10	0.01 to 0.10
Titanium	0.0005 to 0.005	0.01 to 0.10	0.01 to 0.10
Vanadium	0.005 to 0.05	0.001 to 0.01	0.01 to 0.10
Bismuth	0.005 to 0.05		
Strontium	0.005 to 0.05		
Cobalt	0.0005 to 0.005	0.003 to 0.03	0.001 to 0.01
Silver	0.001 to 0.01	0.0005 to 0.005	0.0005 to 0.005
Cadmium	0.001 to 0.01	0.003 to 0.03	0.0005 to 0.005
Potassium	0.30 to 3.0		

Table 17.29

Electric Furnace Dust Composition
(Reference 44)

Component	Weight, %
ZnO	37
Fe	25
CaO	6
MnO	4
Al_2O_3	3
SO_3	3
SiO_2	2
MgO	2
CuO	0.2
P_2O_5	0.2

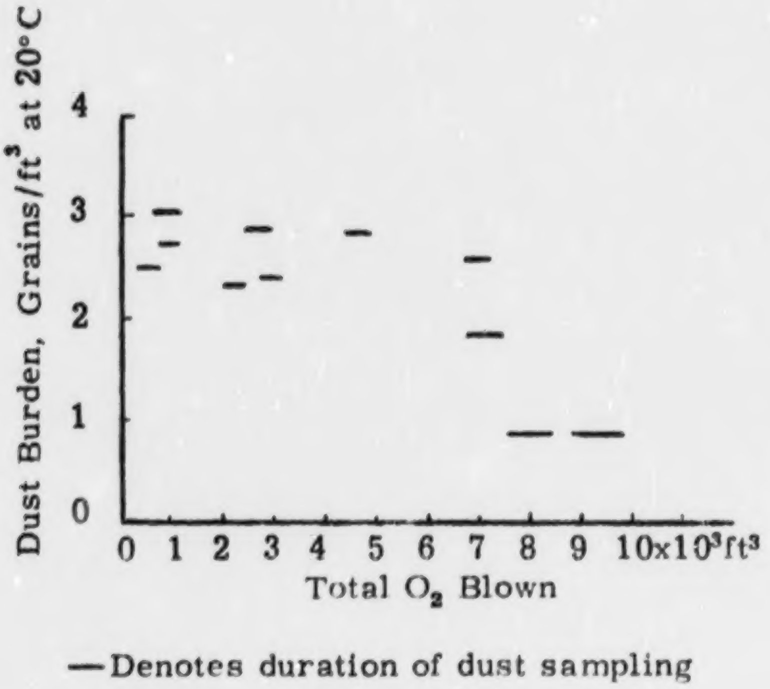

Figure 17.45. Variation of Fume Burdens During Oxygen Lancing.

Table 17.30

Electric Steel Furnace Fume Particle Size Data

Reference	Percent by weight below 0.5 micron	1.0 micron	3.0 microns	5.0 microns
40		90		
54	95		100	
55	90			
56				70
44				71.9

varying operating conditions. Figure 17.46 is an analysis of a rather dense sample of particulate collected from an electric arc furnace.

Ventilation methods for controlling electric arc furnace emissions. The actual amount of gas which must be treated in the dust control system varies substantially with the type of collection equipment utilized. There are basically three types of ventilation techniques used to control effluent from the electric arc furnace. These are: (1) general ventilation, (2) canopy hoods over the furnace, and (3) direct extraction of the fume from the electric arc furnace.

General ventilation to collect effluent from the electric furnace is probably the least efficient, and requires the greatest volume of air. A volume of up to 200 times the volume of oxygen used in the furnace has been reported.[40] A canopy hood over the electric furnace is favorable in some situations because it may be completely independent of the furnace, and equipment around the furnace. A canopy hood arrangement, located 20 to 30 feet above the furnace, is reported to require four or five times the air volume that is required for a direct roof tap, which requires from 60 to 75 times the amount of oxygen lanced.[38] Cross currents through the melt shop materially affect the fume path to the collecting hood, and may result in an inefficient operation.

A direct roof tap appears to be the most efficient method of extracting fumes from the electric arc furnace. The total volume of air to be filtered with this method ranges from 15 to 100 times the volume of oxygen lanced into the furnace, depending on the cooling method. Reviews of the advantages and disadvantages of different methods of fume extraction are given in Ref. 39, 52, 53, and elsewhere.

As shown in Figure 17.44, the collecting duct encloses the three electrodes as closely as possible, and is a part of the swinging furnace roof. In the closed position, this duct matches with the entrance to the flame hood. The flame hood is a brick and steel structure mounted beside each furnace, with a single horizontal reversal of gas flow. Its design and size are dictated by the space available. Water is sprayed into this hood up to a rate of 25 gpm, and is completely vaporized. This serves partially to humidify the gases and quench the flames. Suction fans provide ample draft, except during the more violent periods of bath reactions, when some gases escape. The collected gases pass through uninsulated steel flues to a spray tower where the inlet temperature range is 300 to

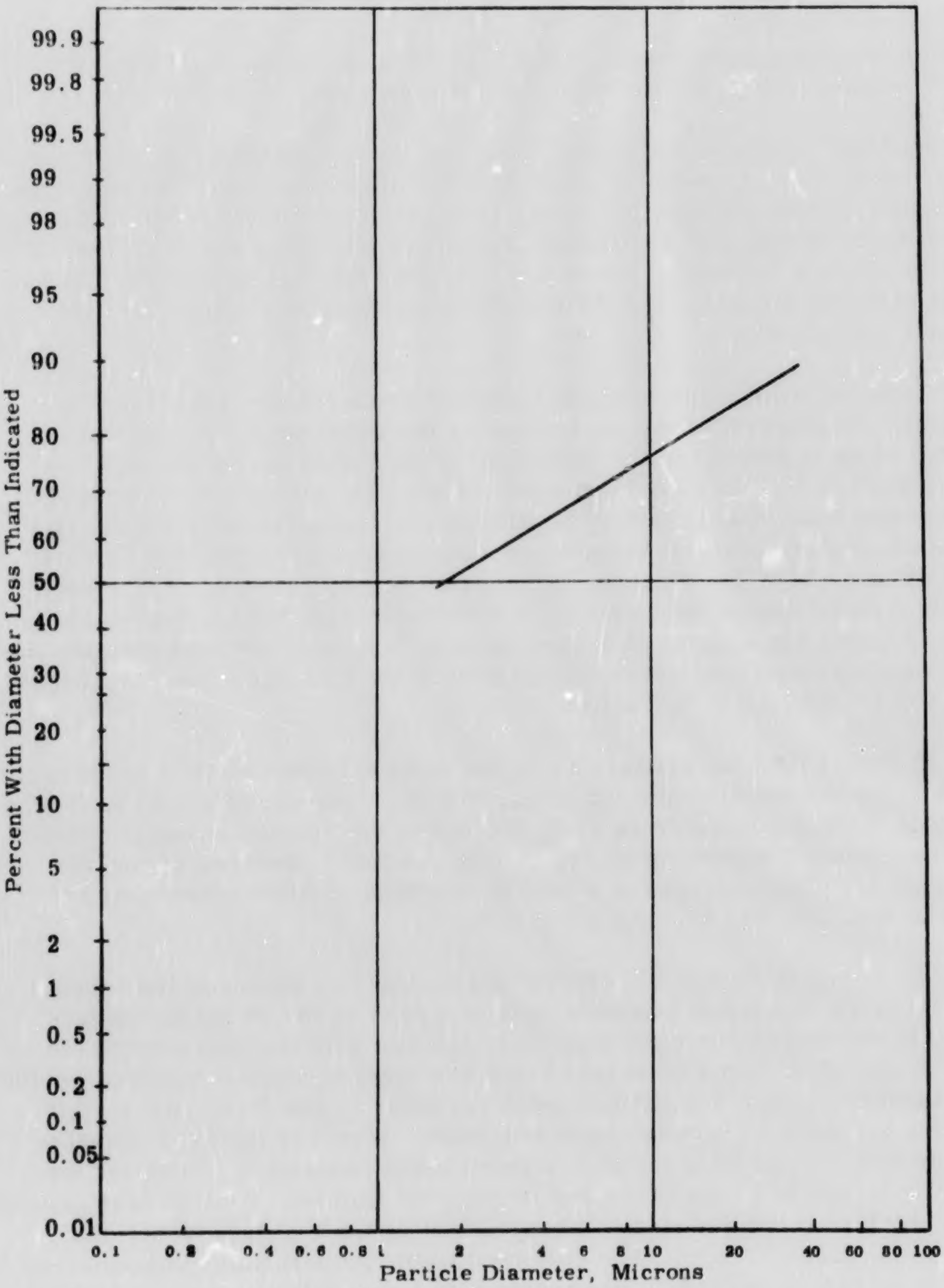

Figure 17.46. Particle Size Distribution By Weight of Electric Furnace Fume. (Reference 47)

500°F, and dust loading up to approximately 1 gr per cu ft.

The gases are cooled by direct contact cooling with water sprays at the first tower, with the quantity of sprayed water controlled by the temperature of the gas in the third tower. Water from the towers is recirculated.

Where machine turnings, oily scrap and bundles, containing a number of non-ferrous metals are charged, it may be necessary to scrub the gases first, and then pass them through an electrical precipitator.

Precipitators for electric arc furnaces. Electrostatic precipitators have been used on several electric furnace installations, both in the United States and Europe. The electrostatic precipitator offers the advantage of a low pressure drop and low power requirements, compared to a bag house or a high energy wet scrubber, and it is capable of achieving relatively high efficiencies.

The precipitator data (Table 17.31, Column A) is for a plate-type electrical precipitator utilizing full-wave rectification.[42] The furnaces are operated under a draft of 0.02 in. water, with the fumes removed through a direct furnace tap into a 2-stage evaporative cooler. The precipitator handles 105,000 cfm at 127°F, with an inlet fume concentration varying from 0.68 to 1.35 grains per cu ft. Collection efficiency of over 97% is realized, using an electrode area to volume ratio of 22.2 sec per ft^2, and a gas velocity of 3.9 ft per sec, giving a gas retention time in the equipment of 4.6 sec.

Another electrostatic precipitator installation (Column B) is operated without any gas conditioning. The electrostatic precipitator equipment is an expanded metal plate precipitator using full-wave rectification. A ring-type hood over the furnace directs the effluent through a radiation cooler, after which tempering air is added to reduce the gas stream temperature to about 80°F. Approximately 33,500 cu ft per min of gas then enters the precipitator with a fume loading of about 0.115 grains per cu ft. The area to volume ratio in this case is 21.5 sec per ft^2, and the gas retention time is 7.8 sec. Gas velocity is 2.3 ft per sec, and a collection efficiency of 92% is obtained. In spite of the more favorable area to volume ratio found in Case B, a lower collection efficiency was obtained, which is probably attributable to lack of gas conditioning.

Table 17.31

Collection of Electric Arc Steel Furnace Fumes
(Reference 20)

Electrical Precipitator Data		
Case	A	B
Operational Data:		
Inlet gas vol., cfm	105,000	33,500
Inlet gas temperature, °F	127	80
Absolute humidity, lb/lb dry gas	0.045	Ambient
Inlet fume conc., gr/ft^3	0.68(1) - 1.35(2)	0.115
Electrical Precipitator Data:		
Type	High eff. plate	Exp. metal plate
Rectification	Mech., full wave	Mech., full wave
Area of collecting surface	38,880 ft^2	11,970 ft^2
Gas velocity, fps	3.9	2.3
Gas retention time, sec	4.6	7.8
A/V ratio, $\frac{\text{Electrode area}}{\text{Volumetric rate}}$, sec/ft	22.2	21.5
Collection Efficiency	97+%	92%
Design Precipitator Rate Parameter	0.16 ft/sec	0.12 ft/sec
Gas Conditioner Data:		
Type	2-stage evaporative cooling	Radiation and tempering air cooling

---

[1] Average for one 50-ton and two 75-ton furnaces processing normal scrap.

[2] Average for one 50-ton and two 75-ton furnaces processing dirty, sub-quality scrap.

Cost data. Capital and operating costs for electrostatic precipitators for electric furnace gas cleaning are given in Table 17.32.[14] These data were taken from the Swindell-Dressler report on costs of control equipment.

17.8 APPLICATION TO SCARFING MACHINES

Scarfing is the operation in the steelmaking process in which the skin of the steel slab is removed. In this operation, slabs from the slabbing mill are conveyed to the stationary scarfing machine, where they are preheated with oxygen and acetylene for about three seconds, after which the scarfing operation begins. The slabs pass the cutting torches at 80 to 120 fpm. A cut of about 1/16 in. is made on two sides of the slab. The sparks and fume are blown downward by compressed air toward a target plate which is continuously sprayed with water.

During the cutting operation, a coarse spray of water is introduced immediately after the gas nozzles to break up the larger chips and wet the fume. The fume is taken through underground ducts to the precipitator where the clean gases are passed through a fan and exhausted to atmosphere through a steel stack. The fume content of these gases (temperature 80-100°F) varies from zero to an estimated maximum of 0.8 grains per cubic foot with an average loading being about 0.1 grain per cubic foot. A horizontal flow duct type precipitator is used in this application. Figure 17.47 is a flow diagram showing the use of precipitators for cleaning scarfing machine gases.

Table 17.33 lists the scarfing machine installations from 1954 to 1969. The design efficiency varies according to the individual plants, but there is no trend shown by the data.

Figure 17.48 is a plot of the cumulative gas volume of scarfing machine precipitators installed over the 1954-1969 period.

Table 17.34 is a listing of average design and performance parameters for scarfing machine precipitators.

Table 17.35 is a listing of capital costs (erected) for two scarfing machine precipitators prepared by Swindell-Dressler. Table 17.36 lists operating costs for two precipitators handling two different gas volumes.

Table 17.32

Electric Arc Furnace - Electrostatic Precipitator[14]

Gas Volume - acfm at 500°F	48,000	185,000	280,000
Furnace Size - Tons	25	150	250
	Capital Costs		
1. Material	$159,000	$465,000	$652,000
2. Labor	85,000	251,000	352,000
3. Central Engineering	61,000	151,000	197,000
4. Client Engineering	15,000	38,000	49,000
TOTAL	$320,000	$905,000	$1,250,000
	Operating Costs		
1. Electric Power	$ 8,000	$30,000	$ 60,000
2. Maintenance	13,000	36,000	50,000
3. Operating Labor	20,000	30,000	40,000
Direct Operating Cost	$41,000	$96,000	$150,000
4. Depreciation	32,000	90,500	125,000
5. Capital Charges	32,000	90,500	125,000
TOTAL	$105,000	$277,000	$400,000

Note: Two-Furnace System

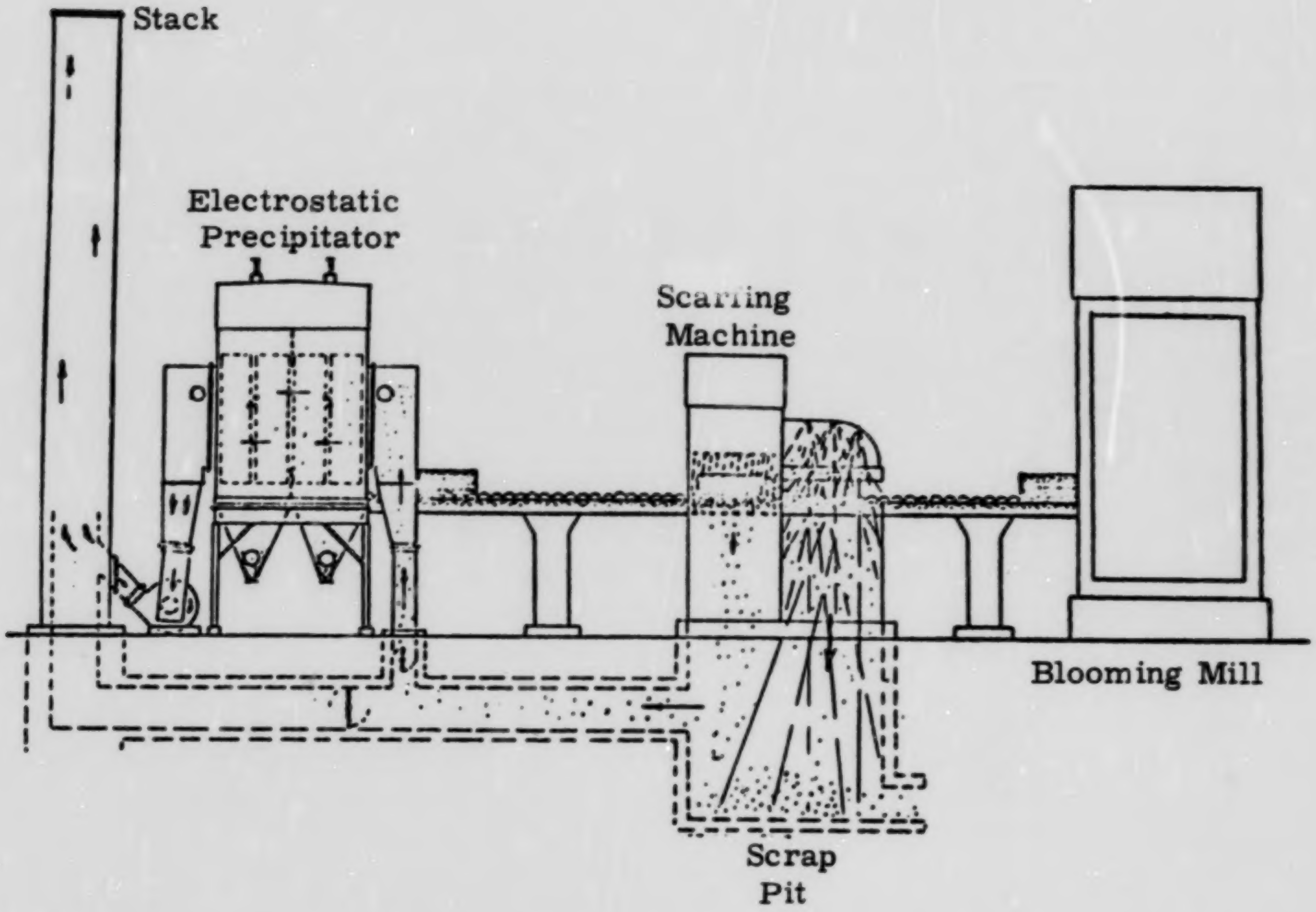
Stack
Electrostatic Precipitator
Scarfing Machine
Blooming Mill
Scrap Pit

Figure 17.47. Flow Diagram Showing Use of Electrostatic Precipitator for Cleaning Scarfing Machine Gas.

Table 17.33

Precipitator Installations on Scarfing Machines
(Period 1954-1969)

Precipitator Contract Year	No. of Installations	No. Electric Precipitators	Total Gas Vol. (10^6 acfm)	Total Gas Vol. Accumulated (10^6 acfm)	Design Efficiency
1954	1	2	0.095	0.095	97.5
1955	1	1	0.100	0.195	95.0
1956	1	1	0.079	0.274	85.0
1957	1	1	0.094	0.368	95.0
1960	1	1	0.100	0.468	95.0
1961	1	1	0.050	0.518	95.0
1962	1	1	0.100	0.618	95.0
1965	1	1	0.060	0.678	93.5
1969	1	1	0.100	0.778	97.5
Grand Total	9	10	0.778		

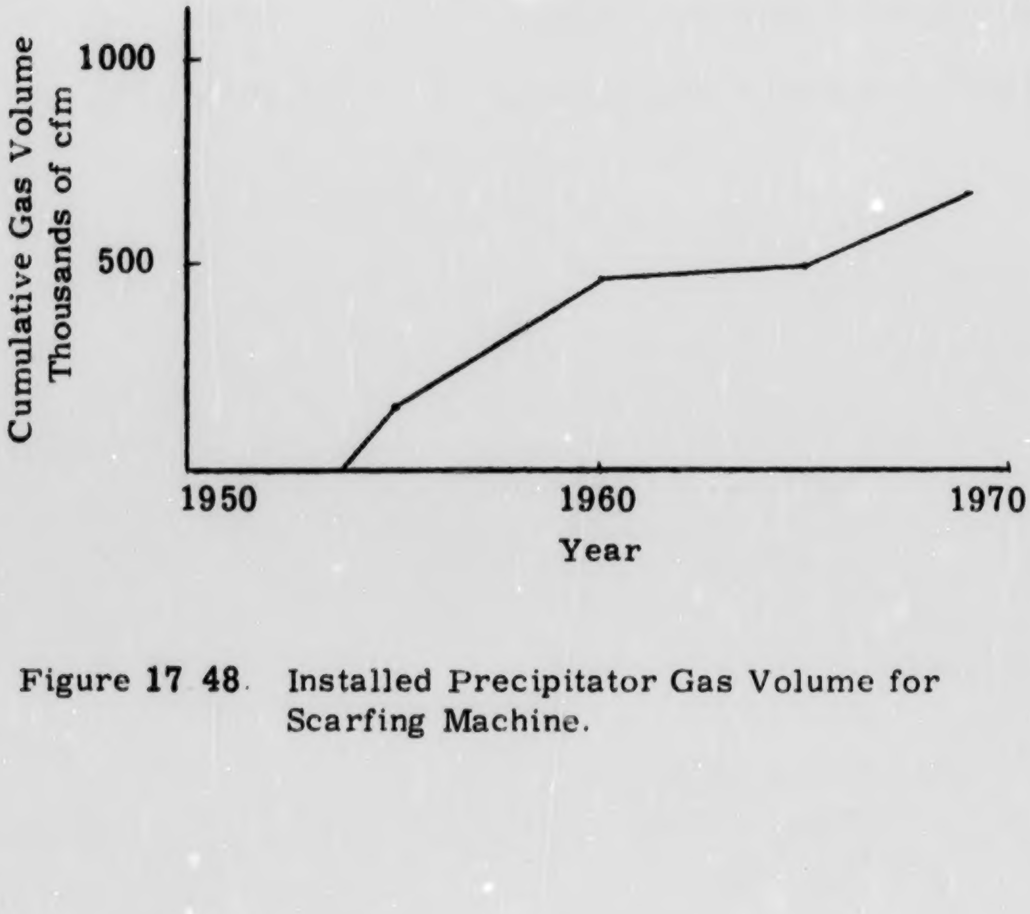

Figure 17 48. Installed Precipitator Gas Volume for Scarfing Machine.

Table 17.34

Design and Performance Data For Electrostatic Precipitators Collecting Dust From Scarfing Machines

No. of Plants	2
Average Gas Velocity	4.5
Average Field Strength (avg. volt./electrode spacing)	11.0
Average Power Input (watts/1000 cfm)	268
Average Precipitator Inlet Gas Temp.	97°F
Average Precipitator Inlet Dust Loading	0.06 gr/scfd

Table 17.35

Electrostatic Precipitators for Scarfing Machines
(Reference 14)

Capital Cost

Gas Volume - acfm at 100°F	50,000	100,000
1. Material	$135,000	$204,000
2. Labor	85,000	112,000
3. Central Engineering	57,000	76,000
4. Client Engineering	14,000	19,000
TOTAL	$291,000	$411,000

Note: Does not include cost of smoke tunnel

Table 17.36

Electrostatic Precipitators for Scarfing Machines
(Reference 14)

Operating Cost ($/Yr.)

Gas Volume - acfm at 100°F	50,000	100,000
1. Electric Power	$ 8,000	$ 18,000
2. Maintenance	12,000	16,000
3. Operating Cost	5,000	7,000
Direct Operating Cost	$25,000	$ 41,000
4. Depreciation	29,000	41,000
5. Capital Charges	29,000	41,000
TOTAL	$83,000	$123,000

17.9 APPLICATION TO IRON CUPOLAS

The iron cupola is a moderately low cost, efficient method of continuously providing molten iron at the desired temperature and chemistry for foundry operations.

The cupola is a refractory-lined cylindrical furnace, somewhat resembling a small blast furnace. When charged with the correct proportions of pig iron, scrap, coke, and flux and when supplied with combustion air, a self-sustaining exothermic reaction takes place due to burning of the coke, which provides the heat necessary to melt the charge and maintain the iron at the desired temperature. Combustion air can be preheated by passing it through heat exchangers which derive their heat from either a separate source of fuel or from the hot exhaust gases emitted by the furnace. Changes in production rate can be obtained by altering the preheat temperature and air blast volume. A cupola of this type is termed a hot blast cupola. A flow diagram for this type of unit is shown in Figure 17.49.

A cupola operating without preheating of the combustion air is termed a cold blast cupola. A flow diagram for this type cupola is shown in Figure 17.50.

Cupola operation is usually cyclic because of the limited life of the refractory linings. If a continuous stream of molten iron is required for long periods of time, more than one furnace must be employed. Typical heat cycles last from eight to sixteen hours.

Considerable quantities of effluent, both gaseous and particulate, are evolved, and the amount is seldom constant except for short periods of time. The effluent rate varies with blast rate, coke consumption, physical properties of coke, type and cleanliness of metal scrap in the charge, coke-to-iron ratio, bed height, burden height, and preheat temperature. The effluent rate also changes at intervals when the furnace is charged with iron, steel scrap, coke, and flux.

Gaseous emissions. Stack gases consist essentially of carbon dioxide, carbon monoxide, sulfur dioxide, nitrogen, and oxygen. Of these, sulfur dioxide is probably the most objectional, both from the standpoint of nuisance odor and as a cause of corrosion damage to equipment. Concentrations of sulfur dioxide over a range of 25 to 250 parts per million by

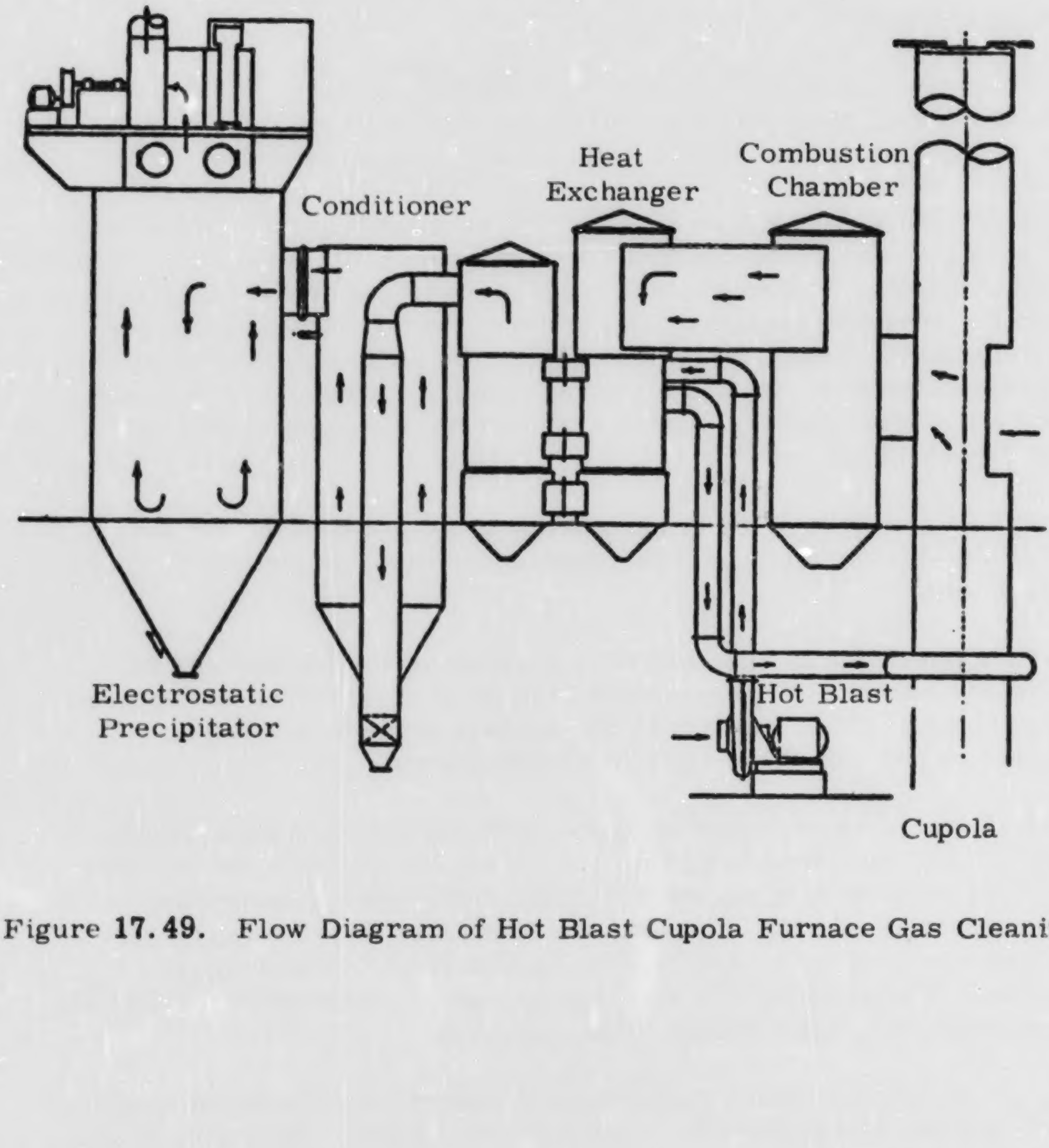
Heat Exchanger
Combustion Chamber
Conditioner
Charge Door
Electrostatic Precipitator
Hot Blast
Cupola

Figure 17.49. Flow Diagram of Hot Blast Cupola Furnace Gas Cleaning System.

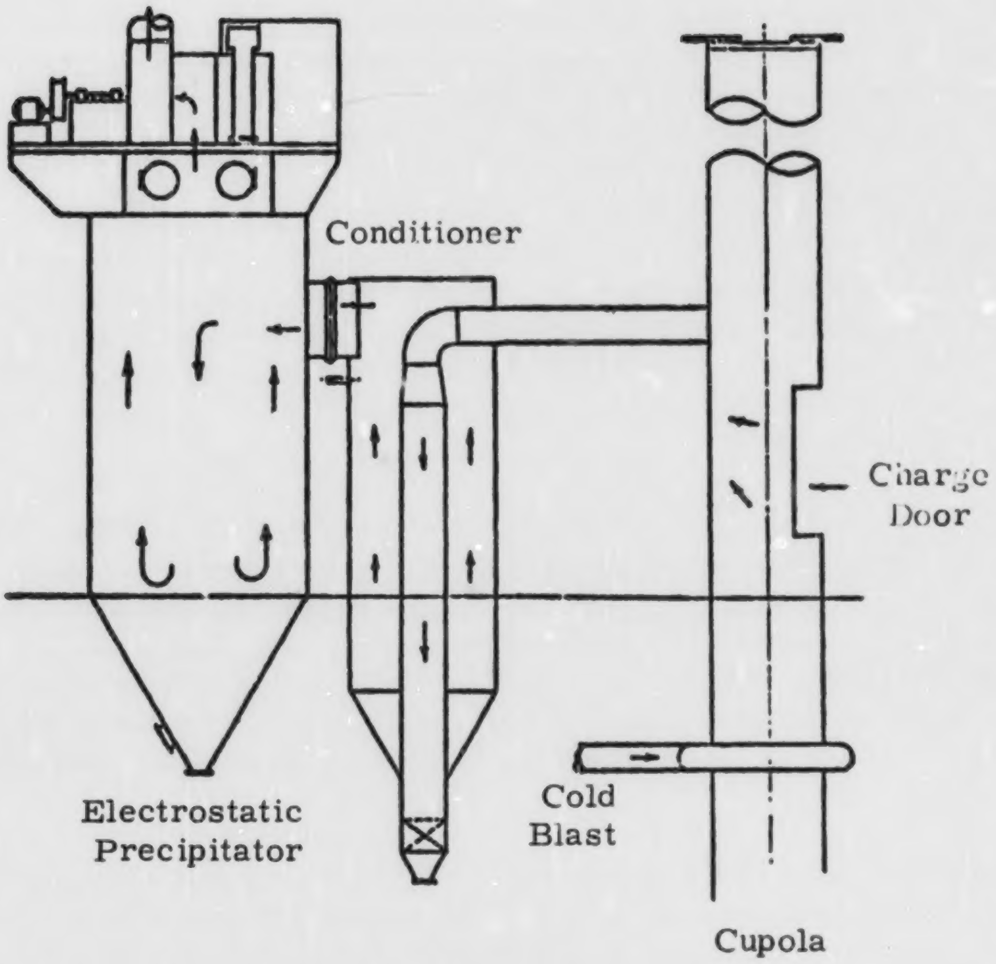
Conditioner
Charge Door
Electrostatic Precipitator
Cold Blast
Cupola

Figure 17. 50. Flow Diagram of Cold Blast Cupola Furnace Gas Cleaning System.

volume have been observed.[57] These concentrations, when combined with moisture, can form enough sulfuric acid to corrode surrounding equipment. Carbon monoxide is ordinarily burned to carbon dioxide in the cupola stack. If it is not burned, it presents a possible explosion hazard in pollution control equipment.

The gases are handled in one of two ways depending upon the type of blast system used:

1. In the cold blast system, the cupola stack is capped and the discharge gases are drawn off and conducted to the cleaning equipment. Temperatures at the cupola exit usually range from 1200° to 2000°F.
2. In the hot blast system, air preheaters are used to preheat the blast air for the cupola. The combustion gases from the cupola furnish the heat for the air preheaters. The cupola stack is capped and the gases are drawn off and sent from the preheaters to the cleaning equipment. Temperatures at the outlet of the preheaters range from 500° to 800°F.

Particulate emissions. Particulate material emitted from the cupola stack consists primarily of the oxides of silicon, iron, calcium, aluminum, magnesium, and manganese, but may also contain small amounts of coke dust, coke ash, limestone, zinc oxide, and smoldering products from paint, grease, oil, rubber, and other combustibles. A scatter range and a mean range of composition of cupola dust is given in Table 17.37.[58] Total emissions may range from 10 to 45 lb of particulate per ton of iron melted.

The dust load in the stack gas also varies widely with scatter values ranging from 0.9-6.5 gr/ft^3 and 1.3-11 grains per std cubic foot for cold blast and hot blast cupolas, respectively. Particulate concentration ranges and scatter values are given in Table 17.38.[58]

The particle size distribution ranges between wide limits, depending on melt rate, coke usage, scrap formulation, and furnace operating variables. A compilation of information from a variety of sources is presented in Figure 17.51 a and b.[59] Under some conditions, over 50% of the dust may be less than 1μ in diameter, whereas in other cases

Table 17.37

Chemical Composition of Cupola Dust

	Mean Range	Scatter Values
SiO_2	20-40%	10 -45%
CaO	3- 6%	2 -18%
Al_2O_3	2- 4%	0.5-25%
MgO	1- 3%	0.5- 5%
FeO (Fe_2O_3, Fe)	12-16%	5 -26%
MnO	1- 2%	0.5- 9%
Ignition Loss (C, S, CO_2)	20-50%	10 -64%

Table 17.38

Dust Content of Cupola Waste Gases

	Mean Range	Scatter Values
Cold blast cupola:		
Top gas, undiluted, grams/Nm^3	6 -11	2 -15
Waste gas from cupola stack, grams/Nm^3	2 - 6	1 - 8
Total dust emission, kg/t iron	5 -10	2 -12
Hot blast cupola (acid)		
Top gas, undiluted, grams/Nm^3	6 -14	3 -25
Waste gas from recuperator, grams/Nm^3	3 - 7	1 -10
Residual gas from cupola stack, grams/Nm^3	0.5- 3	0.2- 7
Total dust emission at cupola furnace, kg/t iron	8-12	2-20

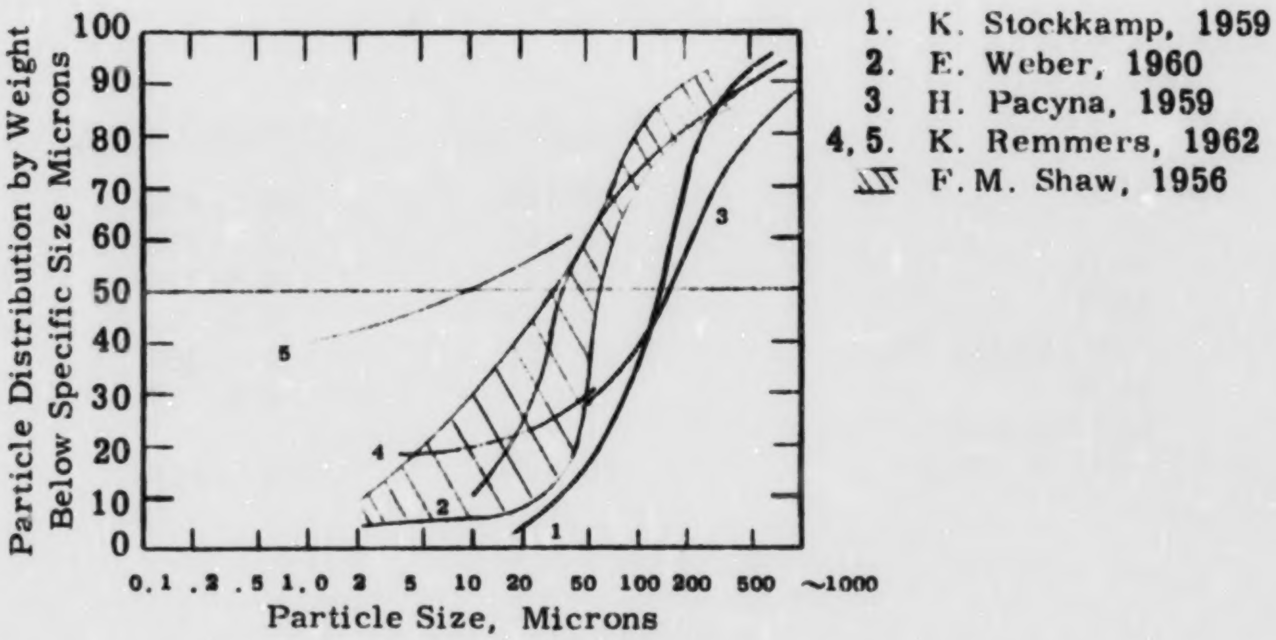

Figure 17.51a. Particle Size Distributions from Cold Blast Cupola Furnaces.

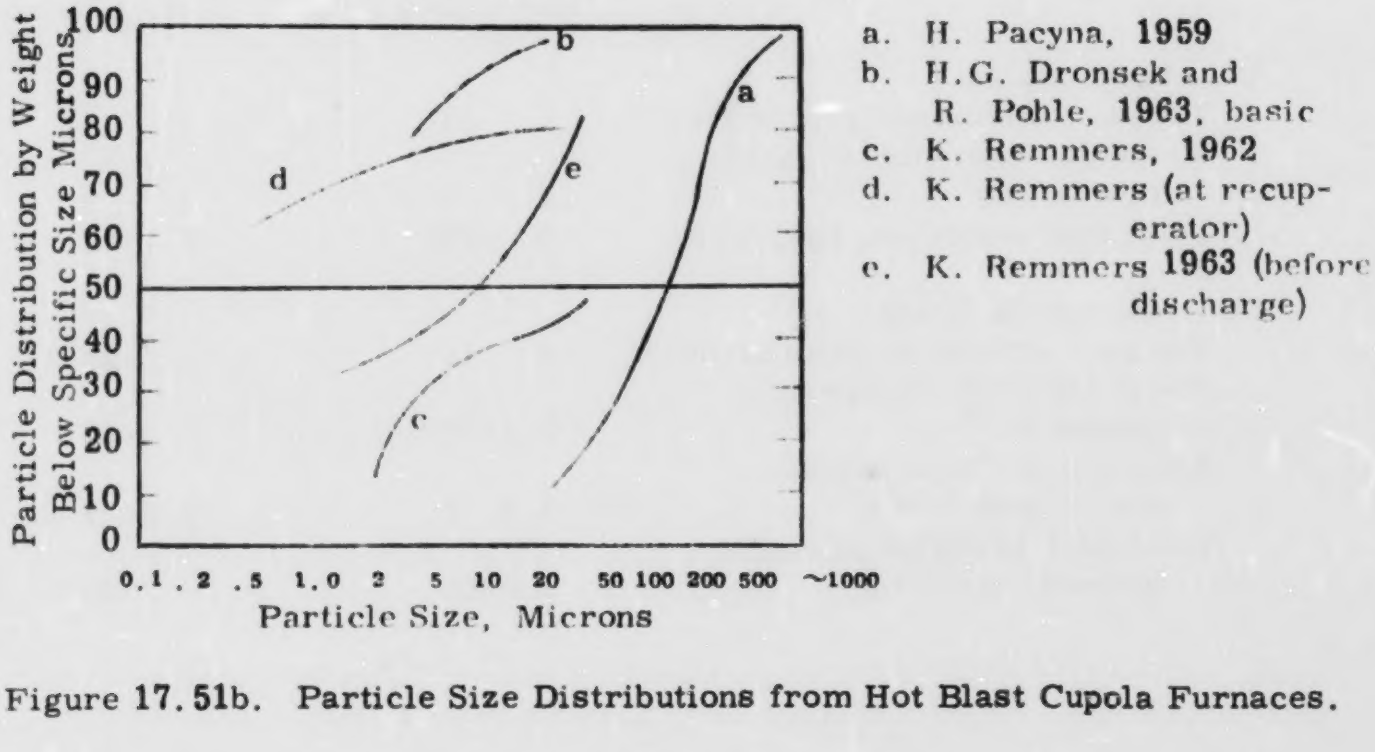

Figure 17.51b. Particle Size Distributions from Hot Blast Cupola Furnaces.

less than 5% by weight of the total particle may be under 2μ in diameter.

It is noted that the older data indicate no more than 10% of the particles by weight are in the size range below 10μ. However, later studies indicate that particle sizes under 1μ may constitute 40% and more of the total weight. Some of the scatter between the various investigations may be a result of differences in sampling equipment and techniques and of sampling at different locations in the ductwork.

There appears to be a relationship between the coke ratio (coke/scrap ratio) and the melting rate per unit area of furnace and the emissions from a cupola stack, as illustrated in Figure 17.52.[59] While a substantial amount of scatter exists, increasing amounts of coke per unit of iron melted in basic hot blast cupolas, generally increase the amount of particulate material. Higher melting rates tend to decrease the amount of particulate when expressed in terms of weight per unit of iron melted. Although the data indicate a wide range of dust emission, of from less than 0.5 to over 30 lb/t iron, most of the data points fall within the range of 4 to 12 lb/t iron.

The effect of blast volume on emissions is illustrated in Figure 17.53.[58] Increasing blast volumes per unit area of furnace cross section increases the lifting forces on dust in the furnace and causes them to be carried upward and out of the stack with the products of combustion.

Particle size distribution data for both hot and cold blast cupolas are presented in Figure 17.54.

Emissions from hot blast cupolas are generally higher than from cold blast cupolas. This is porbably due to larger quantities of small steel scrap being charged into the hot blast cupolas, and is not a direct function of the combustion air temperature, blast volume, or other operating parameters.

Precipitator applications. Application of electrostatic precipitators in this country has been limited to relatively few installations. The variability in cupola emissions, coupled with problems of high resistivity and small particle size, generates problems that have not been completely solved. These same problems also add to the difficulty of proper dust control with other control devices, such as wet scrubbers and fabric filters.

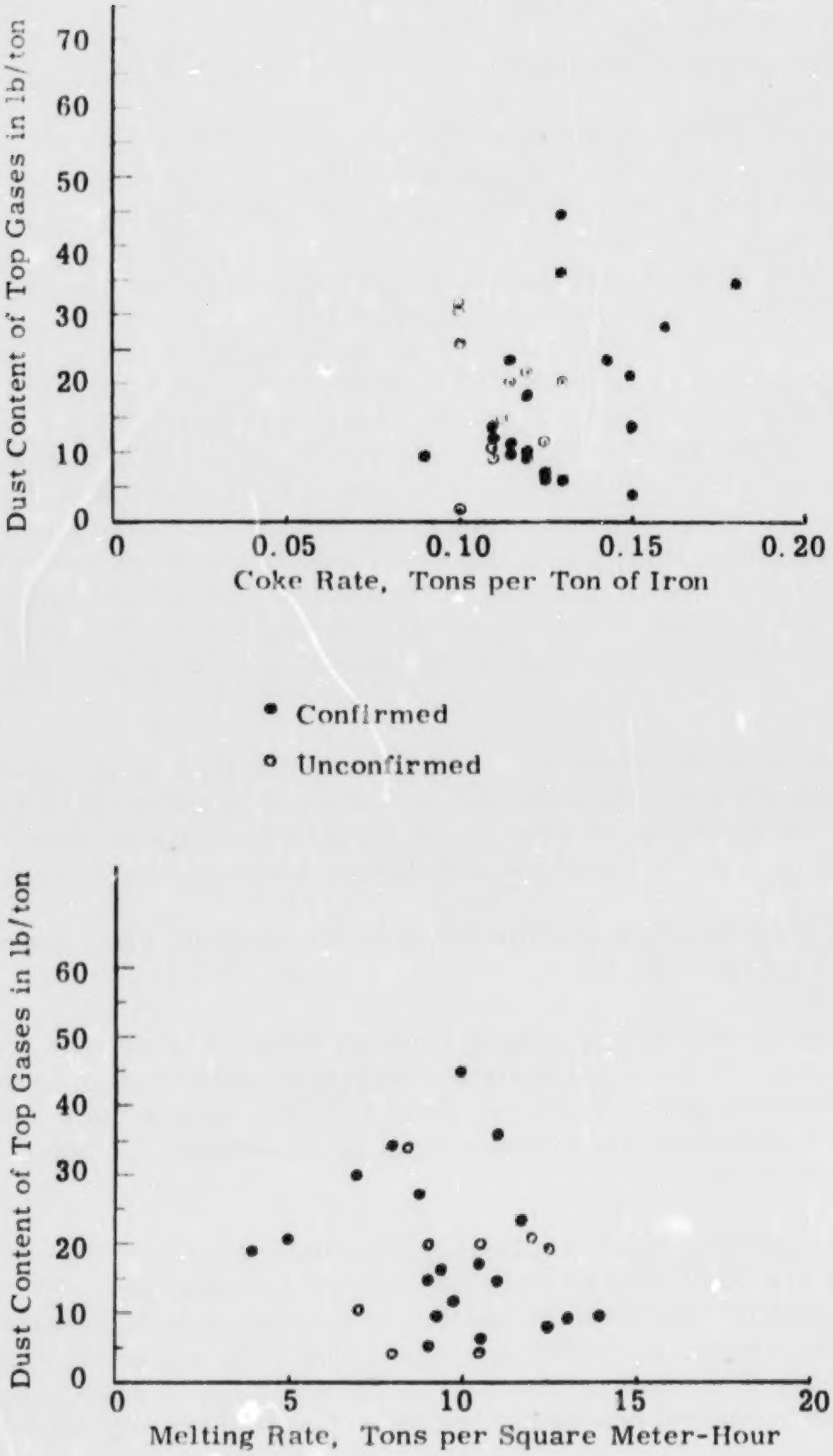

Figure 17.52. Dust Output Survey as a Function of Coke Rate and Melt Rate for Hot Blast Cupolas.

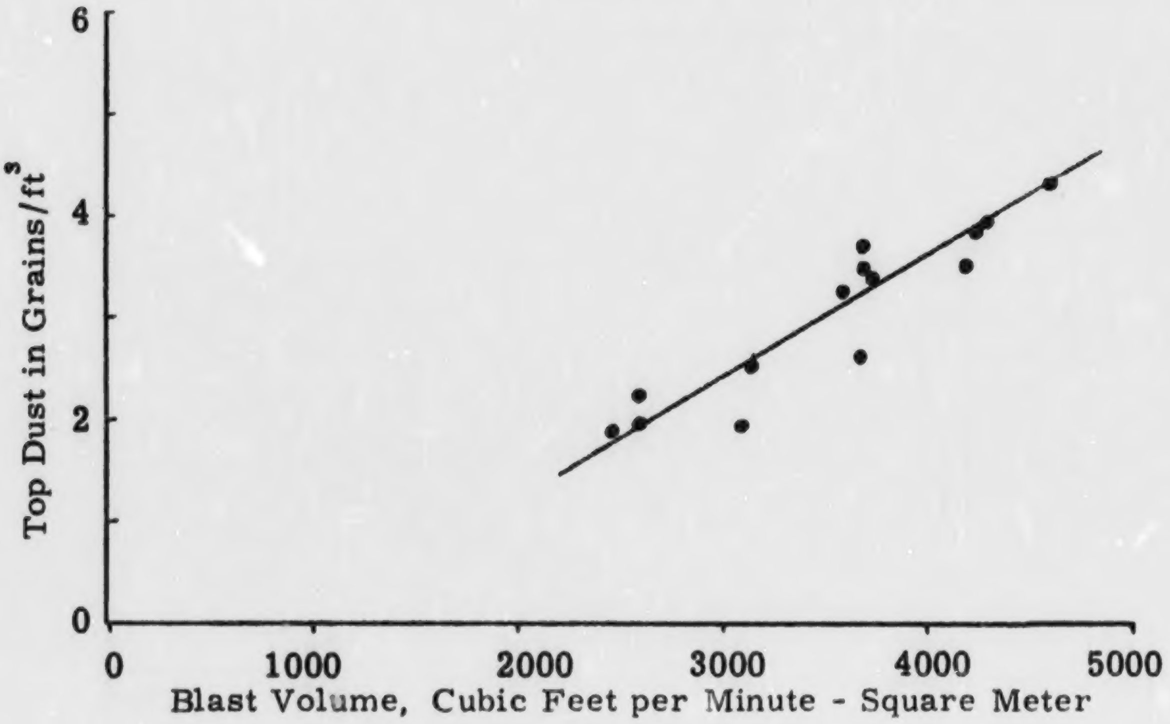

Figure 17.53. Dust Content of Top Gas as a Function of Specific Blast Volume.

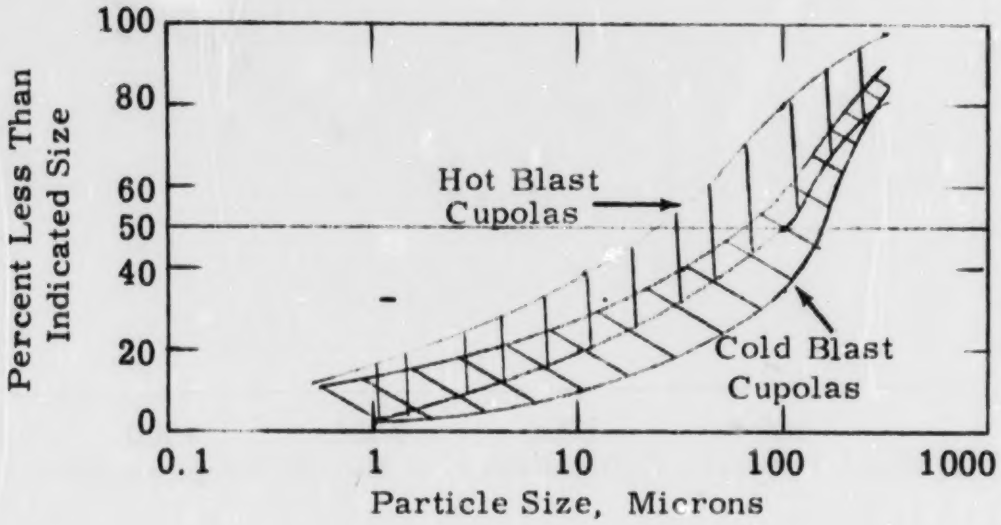

Figure 17.54. Particle Size Ranges for Dusts from Cold and Hot Blast Cupolas.

The installations of electrostatic precipitators on cupolas require that the cupola stack have provision for the introduction of air for burning the CO to CO_2, to prevent fire or explosion in the precipitator.

After combustion, the gases from the cupola must be cooled to reduce the temperature to 300° - 400°F for collection by the precipitatators. The accompanying evaporation of water results in an increase in humidity, which helps to reduce resistivity. The resistivity of dusts from two cupolas is shown in Figure 17.55 as a function of temperature at several moisture contents.[60]

The cooling towers are constructed of steel and equipped with fine sprays to assist in water evaporation. The size of the cooling towers and the quantity of water sprayed must be arranged so that no moisture carry-over to the precipitator occurs. The system's design must be based on the varying gas temperatures and volumes that accompany cupola operation.

Because of the limited number of installations of electrostatic precipitators in foundry cupolas, very little data are available on their performance. The following is a tabulation of some parameters based on a very small sample:

Melt rate	10 tons/hr
Gas volume	17,000-20,000 cfm
Dust loading	0.3-0.7 gr/scf
Efficiency	90-96.6%
Precipitator plate area	18,800 sq ft
Collector material	Stainless steel
Discharge electrode length	9,500 ft
Inlet gas temperature	370°-430°F
Precipitation rate parameter	0.03-0.06 ft/sec

Trends. The most prevalent type of dust control equipment for foundry cupolas has been the high energy wet scrubber because of its compactness and reliability. Fabric filters require more space and higher investment costs. On the other hand, energy requirements for wet scrubbers are high for the removal of the very fine dust from foundry cupolas.

Use of electrostatic precipitators for control of cupola dusts will depend upon finding an economical method of handling the wide variability in operating conditions so that equipment size and cost can be reduced and reliability increased.

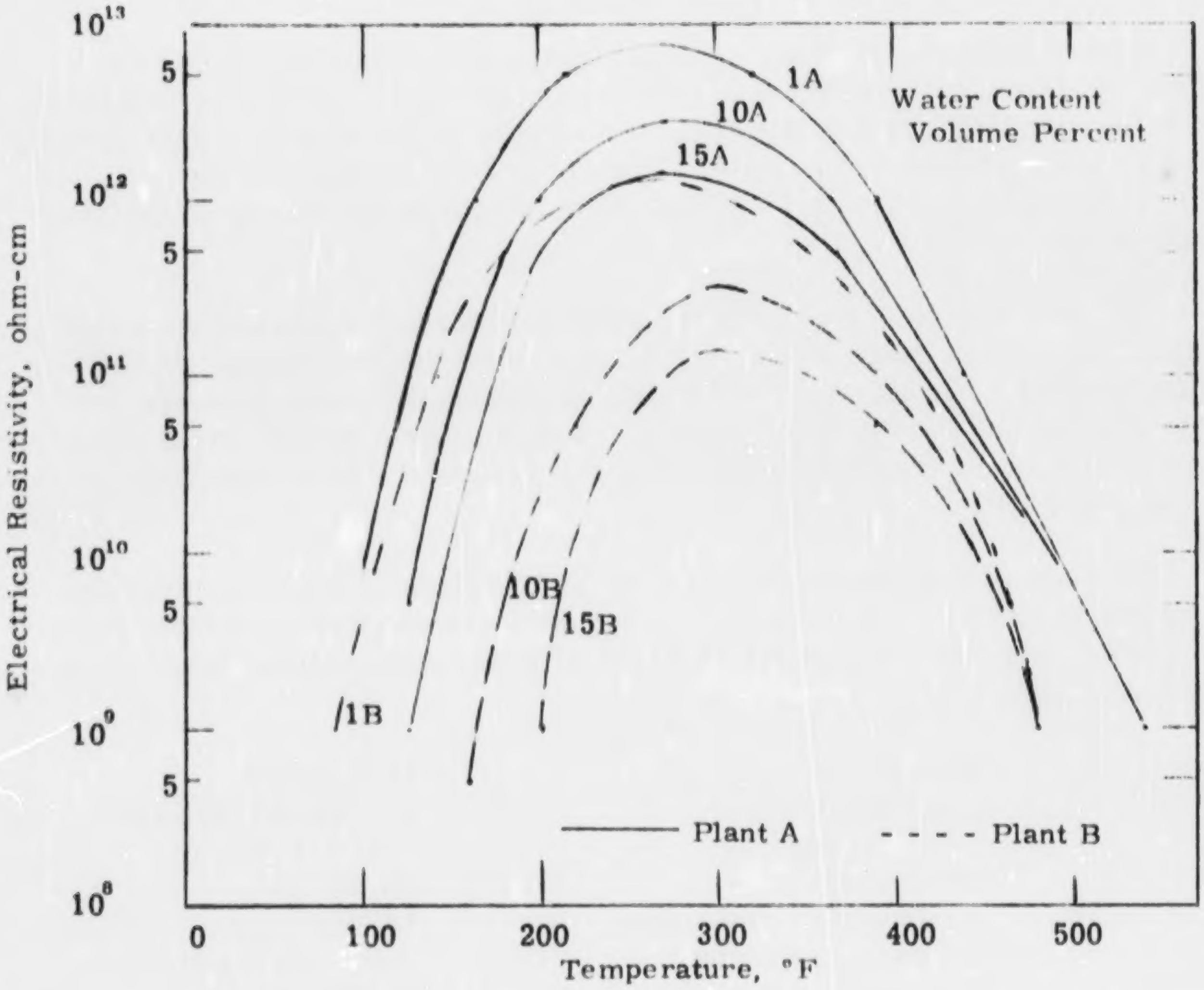

Figure 17. 55. Laboratory Resistivity Measurements for Dusts Collected from Two Cupolas for Various Moisture Contents and Temperatures.

CHAPTER 17
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CHAPTER 18
ELECTROSTATIC PRECIPITATORS IN THE ROCK PRODUCTS INDUSTRY

For purposes of this study, the rock products industry includes the manufacture of portland cement by both the wet and dry processes, gypsum processing, and the various associated raw material and finished product operations requiring dust or fume collecting equipment. The use of electrostatic precipitators in the rock products industry is in the collection of dusts from cement kilns, from gypsum calciners, and from mills and dryers used in preparation of the raw material feed to the kilns and calciners.

Cement is used as an intermediate product for many materials including concrete, mortar, concrete block, precast panels, and concrete pipe. Because of its widespread use in high sales-volume products, the quantity of cement produced annually is very large and has increased rather steadily since its introduction, as shown by the graph in Figure 18.1.

The gypsum industry is one of the most important nonmetallic industries in the country, based on quantities processed. Large tonnages are used as a retarder in cement, as a soil conditioner, and as the main constituent in plaster and prefabricated plaster board. Figure 18.2 shows the production of gypsum from 1935-1969.

18.1 MANUFACTURE OF PORTLAND CEMENT

The raw materials for producing portland cement include lime, silica, and alumina with ferric oxide as a fluxing component. Sources of lime include limestone, cement rock, chalk, marl, shell residues, and blast furnace slag. Sources of silica include clay, shale, slate, cement rock, and blast furnace slag. There are five types of portland cement produced, the classification being determined by the composition of the raw materials and the method of production. The determining limitations apply to the permissible amounts of alumina, ferric oxide, magnesia, and alkalis.

All portland cement is produced in either a wet or dry process, although there are several variations on the types of kilns used and the method of introduction of material into the kilns. Figure 18.3 shows the principal steps in cement production. These steps include quarrying, crushing, grinding, transporting, calcining, regrinding, and packaging. The materials delivered to the primary crusher from the quarry are reduced

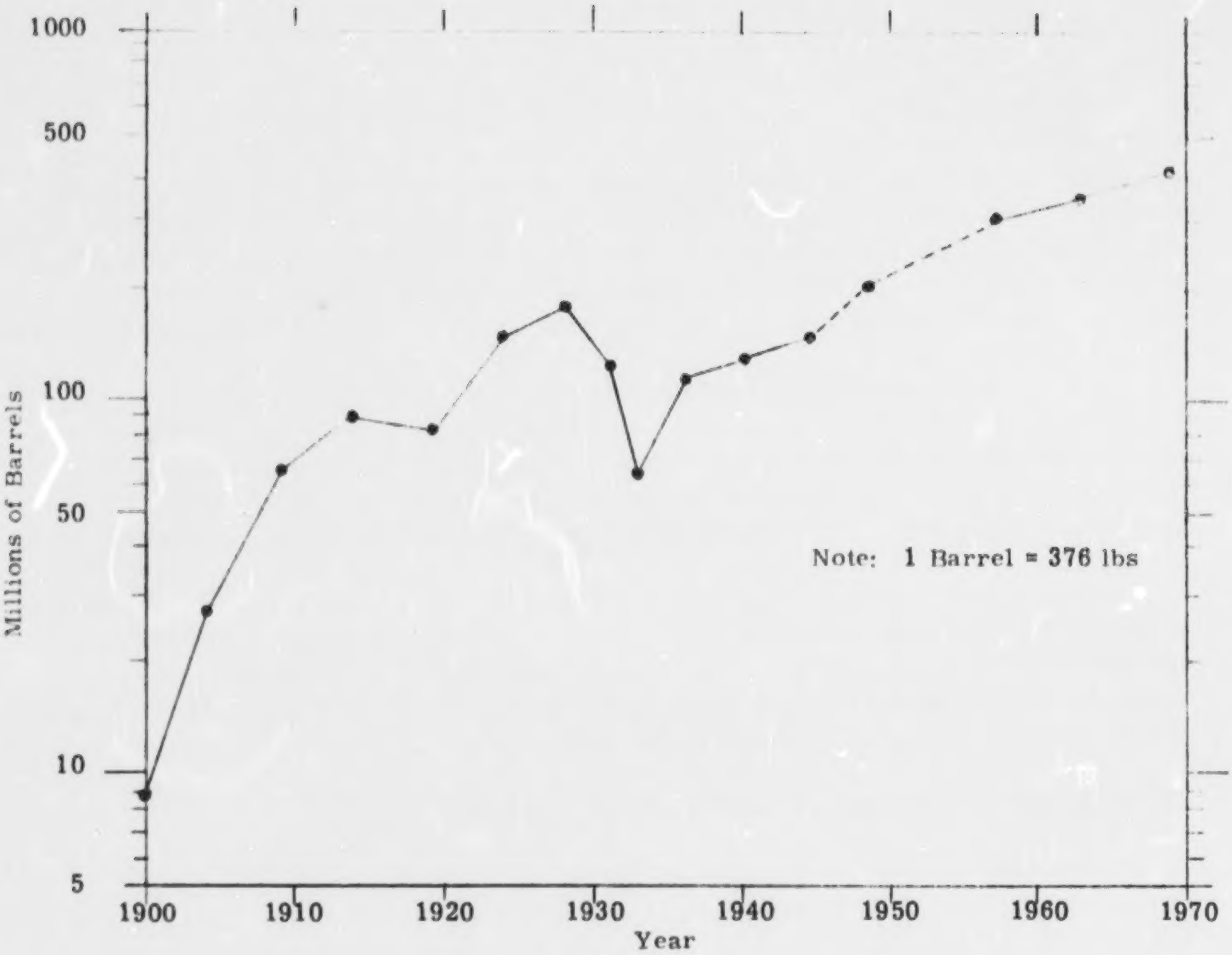

Figure 18.1. Production of Portland Cement in the United States (1900-1969).

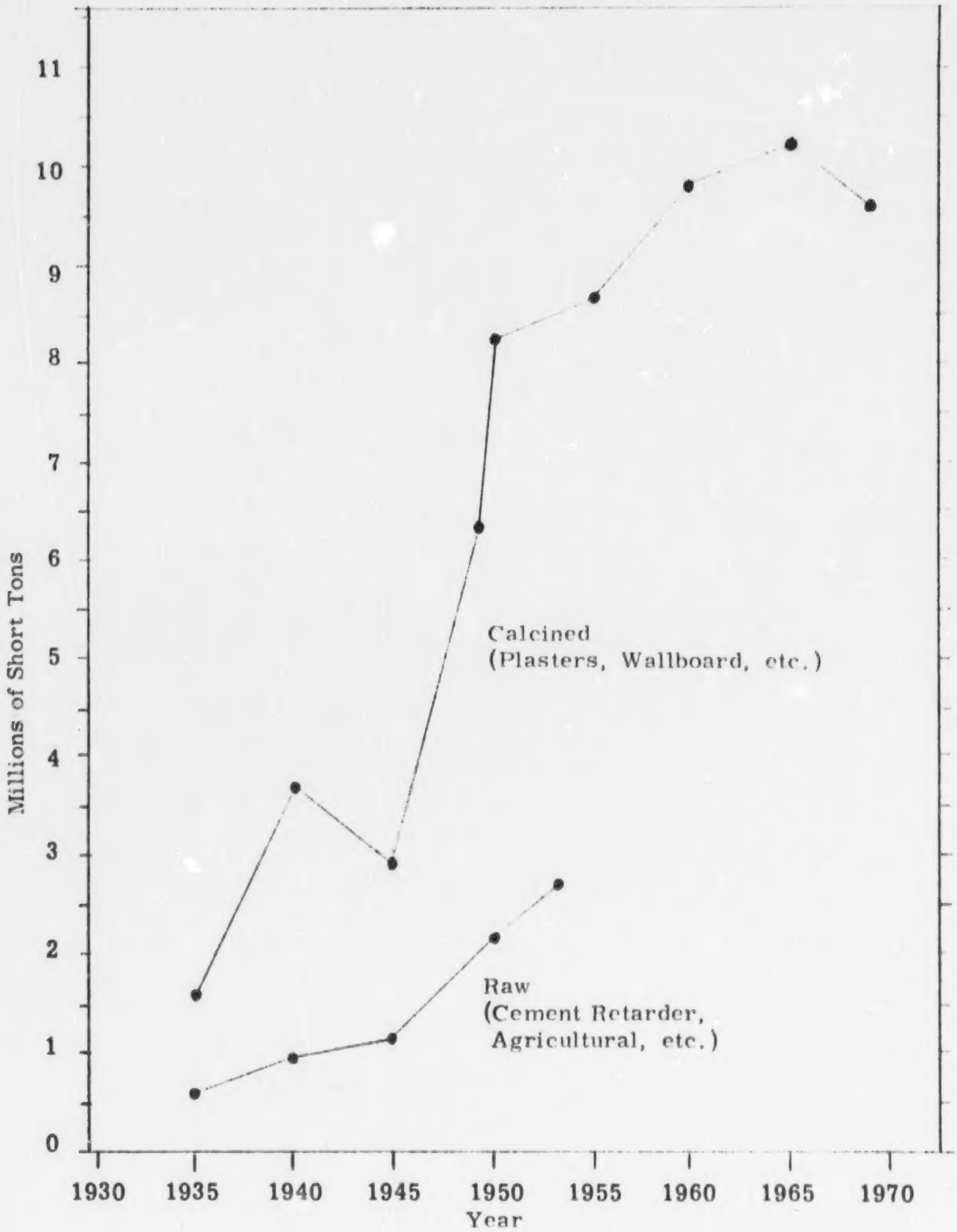

Figure 18.2. Production Statistics of the Gypsum Industry (1935-1969).

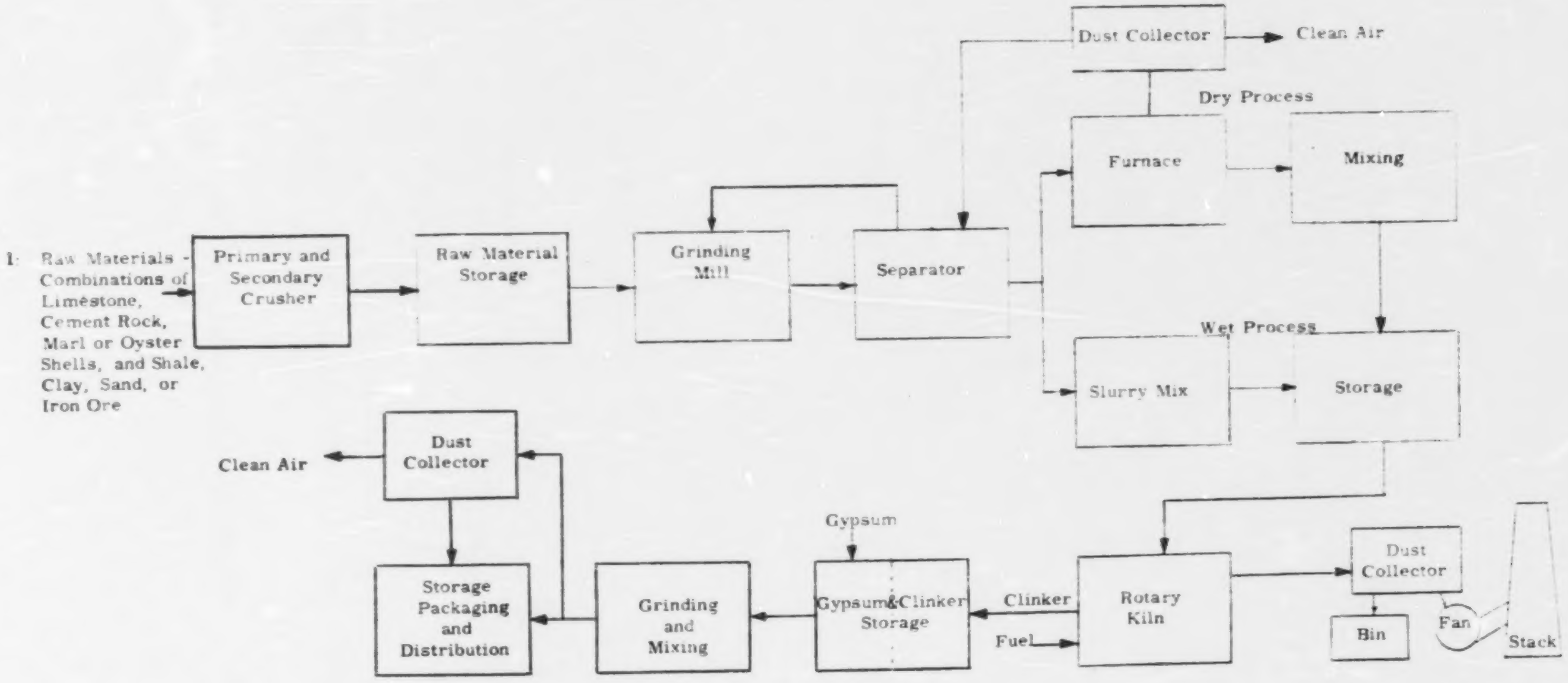
1: Raw Materials - Combinations of Limestone, Cement Rock, Marl or Oyster Shells, and Shale, Clay, Sand, or Iron Ore
Primary and Secondary Crusher
Raw Material Storage
Grinding Mill
Separator
Dust Collector
Clean Air
Dry Process
Furnace
Mixing
Wet Process
Slurry Mix
Storage
Rotary Kiln
Dust Collector
Bin
Fan
Stack
Clinker
Fuel
Gypsum
Gypsum&Clinker Storage
Grinding and Mixing
Storage Packaging and Distribution
Dust Collector
Clean Air

Figure 18.3. Manufacture of Portland Cement. [1]

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[1] Refer to the bibliography for this chapter.

in size from rock that may measure 4-5 feet, to a size range of 6-10 inches. The secondary crusher further reduces the rock to approximately 1/2 to 3/4 inch. From the crusher, the material is sent to a storage bin and then transferred to a wet or dry process grinder for further size reduction and blending. Prior to grinding, the raw materials are proportioned to give the type of cement desired. Grinding mills for both wet and dry process grinding are of the ball, rod, or ring-roller types, and reduce the materials to a size that typically is 70-90% minus 200 mesh. The dry process often requires a special drying operation to reduce the moisture content of the material to below 1% for proper grinding.

The output from the grinders is termed "slurry" in the case of the wet process and "raw meal" in the case of the dry process. Each is fed to the upper end of the kiln, which is the heart of the cement-making operation. The modern cement kiln typically consists of a shell about 11-25 feet in diameter and 500-600 feet long, lined with refractory material. Kilns in the United States are generally erected horizontally with a slope of 3/8 to 3/4 inch per foot, and are rotated about their longitudinal axis.

The material fed to the kiln is heated and dried in the initial section. As it progresses through the kiln it is calcined and heated to a point of incipient fusion at about 2900°F, producing a mineralogical material called clinker. This material, which is ultimately ground into cement, is removed from the kiln at the lower end.

Fuel for cement kilns can be coal, fuel oil, or gas and is burned at the exit end of the kiln with combustion gases passing counterflow to the material. In coal-fueled kilns, the coal is dried and ground to a size of 80-90% minus 200 mesh before injection into the kiln.

The clinker is removed from the kiln at a temperature of about 2800°F and must be cooled prior to final grinding. Clinker coolers typically serve as air preheaters for primary or secondary combustion air.

The final stage of cement production is grinding of the clinker. Grinding is accomplished in ball or tube mills, often preceded by crushers. The final product is ground to 90-100% minus 325 mesh, or an average size of about 10 microns.

The rotary kiln typically requires one million Btu per barrel (376 lb) of cement for a wet process plant and 700,000 Btu per barrel for a dry process plant. Fuel requirements for drying the input feed to the grinding mills for the dry process plant further increase the heat requirements for that type plant. Attempts to reduce the fuel consumption have led to several modifications in the process to recover the heat normally lost. One of the more important of these is the Lepol process which was developed in Europe. The feed in the Lepol kiln is in the form of 3/4 inch pressed pellets of raw meal with 10-12% water added. These pellets are spread on a traveling grate through which the exit gases from the kiln are passed. The gases first pass through a heated portion of the mix and then through the moist pellets. It is reported that the heat requirement for this system is reduced from one million to about 700,000 Btu per barrel.[2] The length of the kiln is also reduced.

The Humboldt preheater is another process variation. The exit kiln gas passes through a number of cyclonic collectors before reaching the dust collectors. The dry kiln feed is heated by the exit gases to about 1375°F before it is charged into the kiln.

The Meig process reduces the water content of the feed slurry from about 30 to 7% in a rotating drum heat exchanger. The exit gas supplies the heat to the heat exchanger.

The Holderbank preheating process utilizes lifters which raise the charge and cascade it through the hot exit gases to utilize the waste heat for heating the charge.

Dust emission problems. The dust emission problems of the cement industry include the minor ones associated with handling, crushing, transporting, and drying of raw material and fuel, as well as the major emission problem associated with the kiln.

Problems with particulate emissions resulting from the fuel are only associated with coal-fueled kilns. In plants utilizing pulverized coal burners, the coal is carried through the grinding process and into the kiln by preheated air. Dust released from the coal-handling system can be handled effectively by fabric filters. Electrostatic precipitators are also used in Europe for the cleanup of dust from coal-handling systems. The principal problems are the design and operation of the facility to minimize the possibility of ignition of a combustible mixture in the precipitator and to rapidly quench such fires as might occur.

A second source of dust emissions is the primary and secondary crushing of the raw materials. The magnitude of the problem depends on the moisture content of the material and on the location of the facility. Where it is a problem, low-temperature fabric filters or mechanical collectors are used.

Conveying and storing operations may also generate dust-control problems. These can be handled by hoods and fabric filters where control is necessary.

Rotary dryers constitute more of a problem in dust control than the process operation described above because of the higher dust loadings and high exit gas temperatures. Sources of particulate emissions from rotary dryers are the fines caused by the abrasion of the material being processed and, in the case of dryers utilizing waste heat, the carry-over from the kiln. Dust concentration from rotary dryers can be expected to be in the range of 5-10 grains per cubic foot. Temperatures may range from 150°F to above 250°F. Mechanical collectors, fabric filters, and electrostatic precipitators are used on this application. For high collection efficiencies, either fabric filters or electrostatic precipitators are used. The moisture content of the effluent from rotary dryers normally gives dust properties favorable to electrostatic precipitation. Problems encountered are variations in emissions caused by varying dryer feed rates, variation in moisture, and corrosion. These problems are similar to those encountered in cleaning of kiln effluents, and are discussed later.

The effluent from finish and grinding mills constitutes another area for dust emission control. The fabric filter has been the primary collector for this area because of the low gas volumes handled and the high electrical resistivity of the dust. The introduction of spray cooling into the mills may increase the moisture to the point that electrical resistivity would be sufficiently low to make electrostatic precipitation more competitive for this application.

18.2 CHARACTER OF EMISSIONS FROM CEMENT KILNS

The effluent from the cement kiln is by far the most difficult problem in gas cleaning for a cement manufacturing operation. The problem is caused by the combination of high gas temperature, high gas volume, and the nature of the particulate. The first source of particulate emission is the dust generated by the grinding and tumbling action within the kiln. Also, the liberation of gases during calcination tends to expel particles into the gas stream. Another source of particulate is the condensation of material that is volatilized during passage through the kiln. The alkalis present in the raw materials fall within this category. The process of vaporization and condensation normally gives rise to particles considerably smaller in size than those produced by mechanical means.

Collection of dust from a cement kiln is important from the standpoint of air pollution control and also for economical plant operation. The effluent from a cement kiln contains large amounts of material of the same composition as the kiln feed, and hence return of this material is of direct economic value. The requirement for air pollution control necessitates higher collection efficiencies than would be indicated by the economics of product recovery alone. However, the return of a portion of the collected dust does help defray the cost of dust control equipment.

One problem common to many cement manufacturing operations is that raw materials fed to the kiln contain more alkali than is permitted in the finished clinker. Most of the alkalis present volatilize at the kiln temperature and are removed with the exit kiln gases. On cooling, condensation of the vapor occurs, producing an exceedingly fine particle or fume.

If all of the effluent from the kiln were collected and returned to the kiln, the alkali content of the product would soon reach an unacceptable level, and this places a restriction on the amount of collected dust that can be returned to the kiln.

One advantage of an electrostatic precipitator is that the smaller particle sizes are more difficult to collect, and a higher percentage of them appear in the final stages of the precipitator. Consequently, there is a fractionating effect tending to separate the alkali-bearing constituent of the dust from the cement-making constituents. This effect permits a higher percentage dust return to the kiln while maintaining the alkali content within acceptable limits. The alkali-rich dust is often disposed of in abandoned quarries, or dumped.

The volume of gases emitted from the kiln ranges from around 45,000 to over 100,000 cfm, depending on the method of preheating of raw materials and the method of gas cooling employed.

Particle properties. Approximately 85% of the dust particles in the exit gas from the kiln are minus 20 microns. These particles have the approximate composition: CaO - 41%; SiO_2 - 19%; Fe_2O_3 - 10%; K_2O - 7%; MgO - 2%; Na_2O - 1%; volatile or combustible - 13%. Figure 18.4 is a size distribution curve for a typical cement kiln dust.

The particulate emission rate for a cement kiln is highly variable because of variations in the raw feed and kiln design. Dust loadings of from 1 to 70 grains per cubic foot have been reported for both wet and dry process plants. Table 18.1 shows the inlet dust loadings reported for 13 wet process and 5 dry process plants. With such wide variation, it is difficult to establish inlet dust load conditions.

The graph, Figure 18.5, shows the relationship between resistivity, temperatures, and moisture content. For good collection efficiency, resistivity of the dust must be maintained in the neighborhood of 10^{10} ohm-cm or lower. If the temperature is between 500° and 600°F, the volume resistivity of the dust is low enough to be within a tolerable range for precipitation, regardless of the moisture content. Below 550°F, the resistivity is extremely dependent on moisture content and it is necessary to have moisture contents in the range of 10-20% to get resistivities within the range best handled by a precipitator.

The gases from cement kilns are composed of nitrogen, water vapor, carbon dioxide, and small concentrations of oxygen, sulfur dioxide, and inert gases. The relative amounts of these constituents depend largely on the type of fuel used. Compositions of the gases by volume are typically as follows: CO_2 - 17-25%; O_2 - 1-4%; CO - 0-2%; N_2 - 75-80%. The moisture content of dry process gas is low, in the range of 1-8% by volume, whereas the wet process has a moisture content in the neighborhood of 30% by volume.

The temperature at which a cement kiln precipitator operates depends upon economic considerations, the alkali content of the kiln gases, air pollution control codes, and the extent of heat recovery used. On dry process kilns where gas cooling by water spray is used, the size of the cooling chamber, and hence the cost, is related to the gas temperature. From this standpoint, the higher the temperature of the gas to the precipitator, the smaller the size of the cooling tower. Operation at higher temperatures also reduces the resistivity problem. On the other hand, lower gas temperatures may be required to insure condensation of the alkali compounds in the gas.

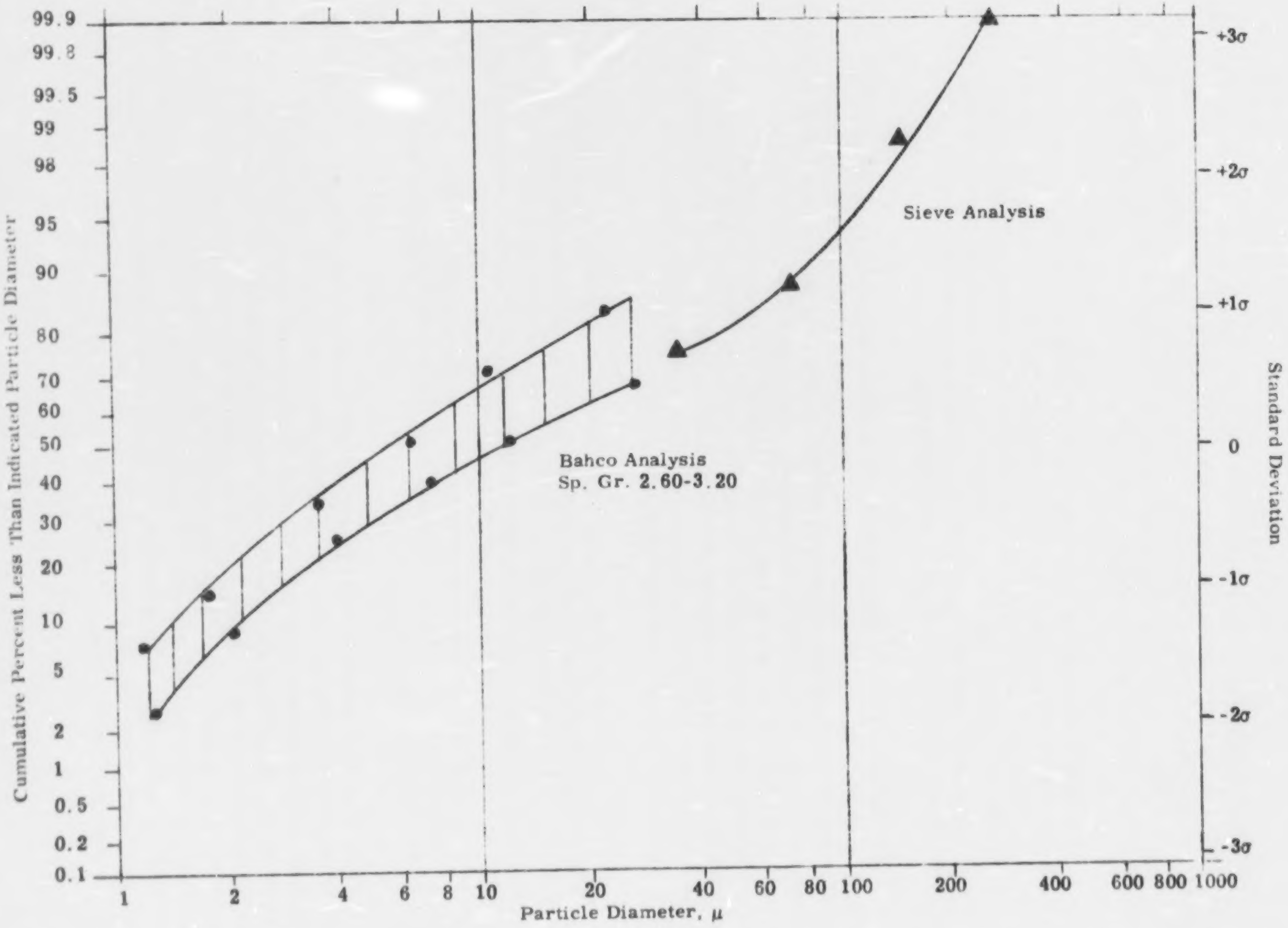

Figure 18.4. Typical Particle Size Range of Cement Dust.

Table 18.1

Precipitator Inlet Dust Loadings for Cement Industry Precipitators

Precipitator Inlet Dust Loading (gr/scfd)	Wet Process		Dry Process		Combination Mills, Dryers, Vents	
	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm
0 - 10	4	0.670	1	0.185	1	0.178
10 - 20	4	0.562	2	0.138		
20 - 30	4	0.151	1	0.254	1	0.049
30 - 40	2	0.417				
40 - 50	1	0.369			1	0.014
50 - 60			1	0.111		
60 - 70	1	0.183				
Totals	13	2.352	5	0.688	3	0.241

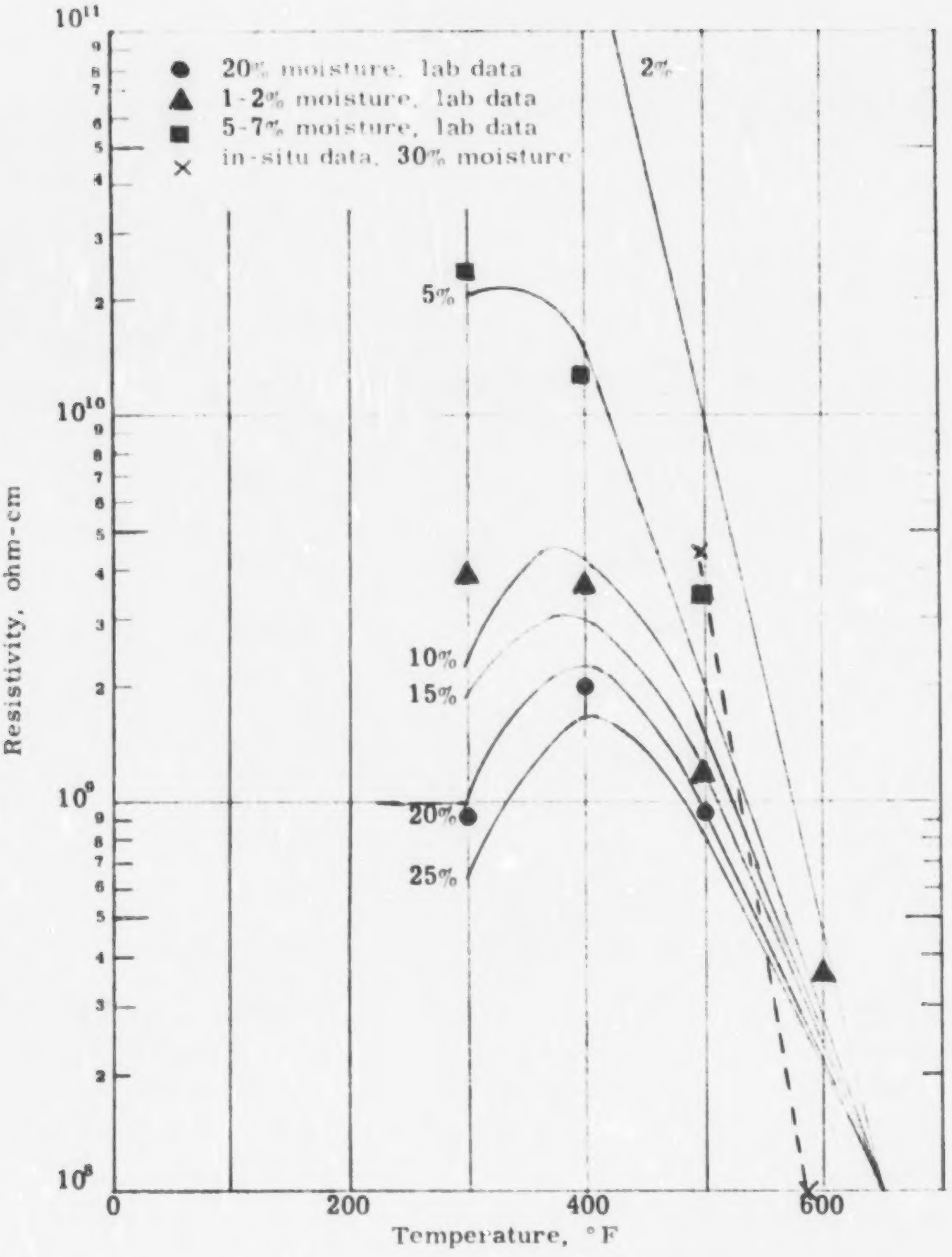

Figure 18.5. Relationship Between Resistivity, Temperature, and Moisture Content for Cement Kiln Dust.

If heat recovery is used, such as in the Lepol process, it is desirable to maintain exit gas temperature high so that the heat content of the cleaned gases can be used for drying or other process heating.

Dust from a cement kiln has high electrical resistivity in the absence of moisture, and can be difficult to collect. In the wet process, moisture in the kiln gases is inherently high and resistivity is less of a problem. However, in the dry process, unless the precipitator is operated at high temperatures, moisture must be added in the process of reducing the gas temperature by the evaporation of water in the spray cooling chamber. Table 18.2 shows the range of moisture content of the gases from wet process and dry process kilns.

Figure 18.6 is a psychrometric chart that is useful for illustrating the effects of cooling kiln gases by water spray. If the exit gas from the kiln is at a temperature of 1300°F and contains about 10% moisture as indicated by Point I, spraying of water into this hot gas will cause simultaneous lowering of temperature and increase of moisture content described by the adiabatic cooling line shown as the upper dotted line in this figure. The moisture content at a temperature of 500°F is 0.25 lb/lb dry air. Similarly, starting with a 3% moisture in the gas and cooling to 500°F would result in a moisture increase to 0.19 lb/lb dry air.

The size of the precipitator required to achieve a given collection efficiency is related to the gas volume handled and the resistivity of the collected dust. Hence, one is constrained to operate in a temperature range of 500-600°F if the moisture content is low. For high moisture content, temperatures of around 300°F yield resistivities in the range that can easily be handled by electrostatic precipitators. Table 18.3 lists the number of precipitators and the inlet gas volumes handled within the indicated temperature intervals for wet and dry process plants and for combination mills, dryers, and vents.

18.3 ELECTROSTATIC PRECIPITATORS FOR CONTROL OF CEMENT DUSTS

Development of electrostatic precipitators for the cement industry began with the installation at the Crestmore plant of the Riverside Cement Company in California in 1911. The collection electrodes were of heavy wire screen with vertical ribs for reinforcement and baffling. Weighted

Table 18.2

Range of Gas Moisture Contents for Precipitators in the Cement Industry

Gas Humidity (% by volume)	Wet Process Number of Installations	Wet Process 10^6 acfm	Dry Process Number of Installations	Dry Process 10^6 acfm	Combination Mills, Dryers, Vents Number of Installations	Combination Mills, Dryers, Vents 10^6 acfm
0 - 10			1	0.111	3	0.241
10 - 20	1	0.307	3	0.392		
20 - 30						
30 - 40	8	1.281				
40 - 50	2	0.313				
Total	11	1.901	4	0.503	3	0.241

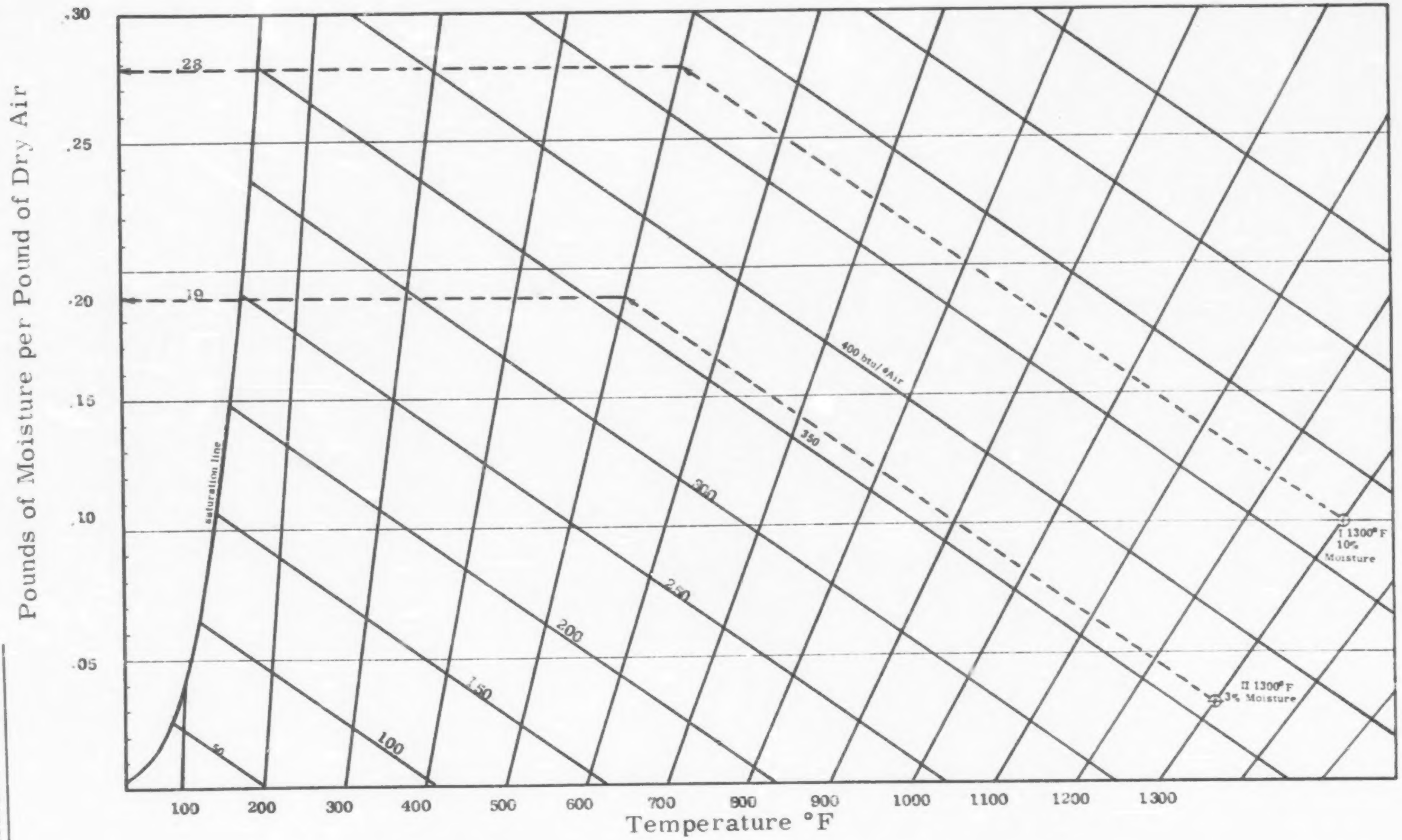

Figure 18. 6. High Temperature Psychrometric Chart.

Table 18.3

Precipitator Inlet Gas Temperatures

Interval Pptr. Gas Temperature (°F)	Wet Process		Dry Process		Combination Mills, Dryers, Vents	
	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm
0 - 100					1	0.014
100 - 200					2	0.227
200 - 300						
300 - 400	1	0.162				
400 - 500	4	0.719				
500 - 600	5	1.189				
600 - 700	2	0.282	2	0.332		
Totals	13	2.352	2	0.332	3	0.241

wires were used as discharge electrodes. Mechanical rectifiers were used in conjunction with a motor generator set and a high-voltage transformer to supply electrical power. Corona wire rapping was carried out manually using hammers actuated through mechanical linkage, while collecting electrode cleaning was by air-actuated hammers.

Some early precipitators for cement kilns used an electrode consisting of grounded reinforcing rods imbedded vertically in the concrete and spaced so that the rods were equidistant from two adjacent wires. Economic factors and fabrication difficulties led to abandonment of this type of collecting electrode in favor of simpler electrode configurations.

Electrostatic precipitators used in the cement industry are almost universally the single-stage, horizontal-flow, duct type with insulated steel shells. Electrodes are of the plate or screen type with fins or baffles for stiffening and prevention of dust erosion.

One of the first applications of electrostatic precipitators in the cement industry was in the recovery of dusts from cement kilns, although in recent years precipitators have also been utilized to recover dust from ventilating air from raw and finish mill grinding operations, rotary dryers, etc.

Table 18.4 lists the number of electrostatic precipitator installations in cement plants from 1939 to 1969. The table also gives the total volume of gas handled by the precipitators installed during the year and the total cumulative gas volume. The average design efficiency on a weighted cfm basis is also given. Figure 18.7 is a plot of the cumulative gas volume handled by precipitators in the cement industry for the period 1939-1969. Figure 18.8 shows the design efficiency trend over the same period. Table 18.5 shows the design efficiency trends for precipitators installed on wet and dry process kilns, and combination mills, dryers, and vents, for the period 1949-1969.

Tables 18.6 to 18.8 show the range of gas velocities, power inputs, and field strengths for precipitators installed in cement plants. Table 18.9 is a summary of performance statistics showing the range of some input and performance variables. The precipitator performance ratio "R" is the ratio of the precipitation rate parameter achieved during tests, to the design value, for a group of 12 wet process, 5 dry process, and 3 combination units, and is some indication of the variation in design parameters.

Table 18.4

Electrostatic Precipitator Installations in Cement Plants
Period 1912 through 1969

Pptr. Contract Year	No. of Install.	No. of Pptrs.	Total Gas Volume 10^6 acfm	Total Gas Volume Accumulative 10^6 acfm	Weighted Design Efficiency on acfm Basis - %
Prior 1939	25	41	4.98	4.98	95.2
1939 - 1951	56	96	13.2	18.18	95.5[1]
1952	3	4	0.22	18.40	96.0
1953	5	10	1.15	19.55	95.6
1954	3	4	0.29	19.84	97.9
1955	22	34	5.17	25.01	97.3
1956	12	21	3.64	28.65	93.2
1957	7	9	0.80	29.45	97.4
1958	4	5	0.76	30.21	98.4
1959	5	5	0.75	30.96	98.2
1960	5	8	1.15	32.11	97.6
1961	2	2	0.33	32.44	98.3
1962	7	12	1.52	33.96	97.9
1963	8	10	1.93	35.89	97.7
1964	2	3	0.49	36.38	99.0
1965	5	11	1.77	38.15	99.7
1966	6	11	1.64	39.79	99.6
1967	1	1	0.27	40.06	99.8
1968	6	12	1.44	41.50	99.7
1969	1	1	0.07	41.57	99.5
Totals	185	300	41.57		

[1] Estimated

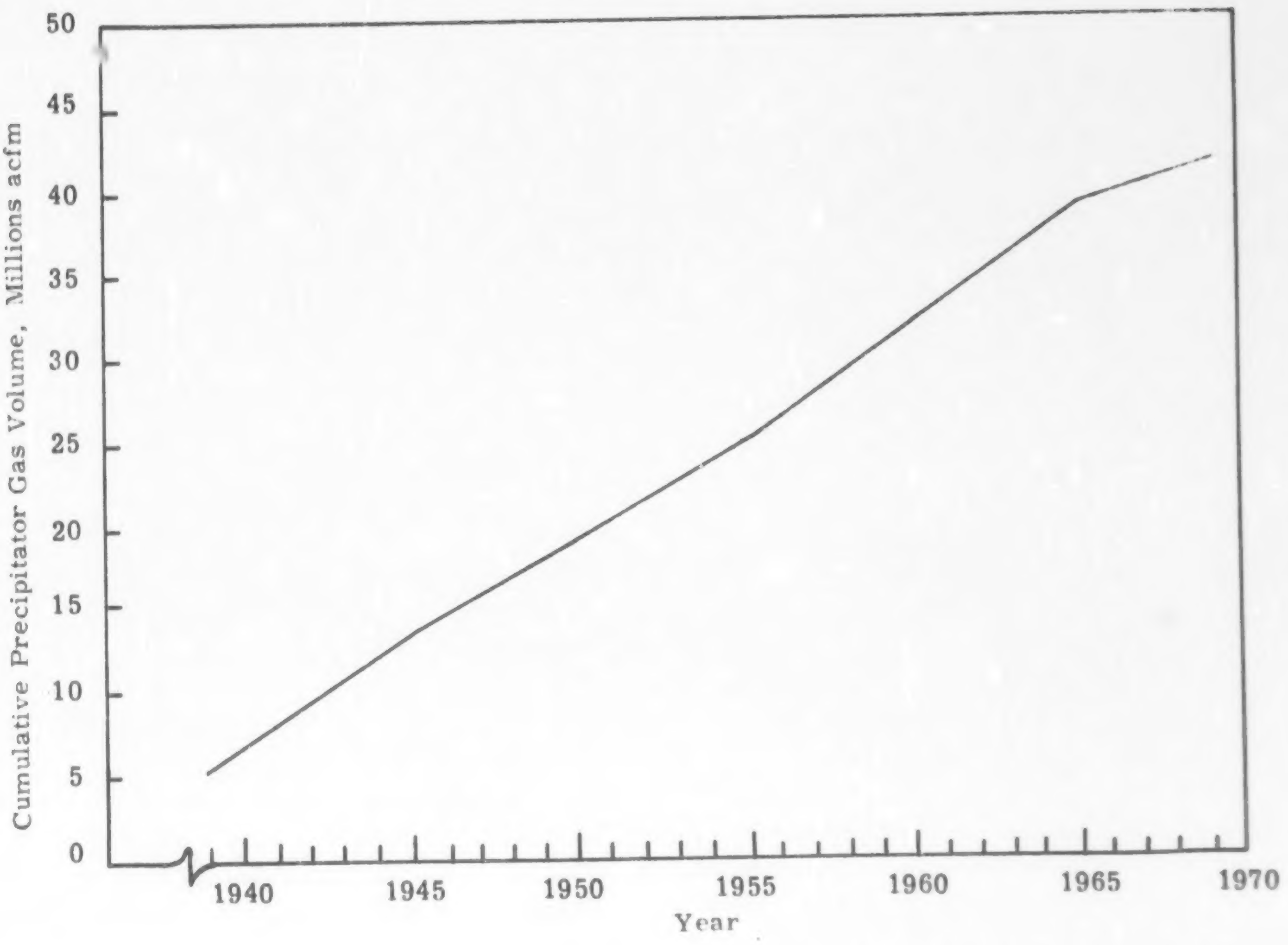

Figure 18.7. Installed Precipitator Gas (Cumulative) for the Cement Industry.

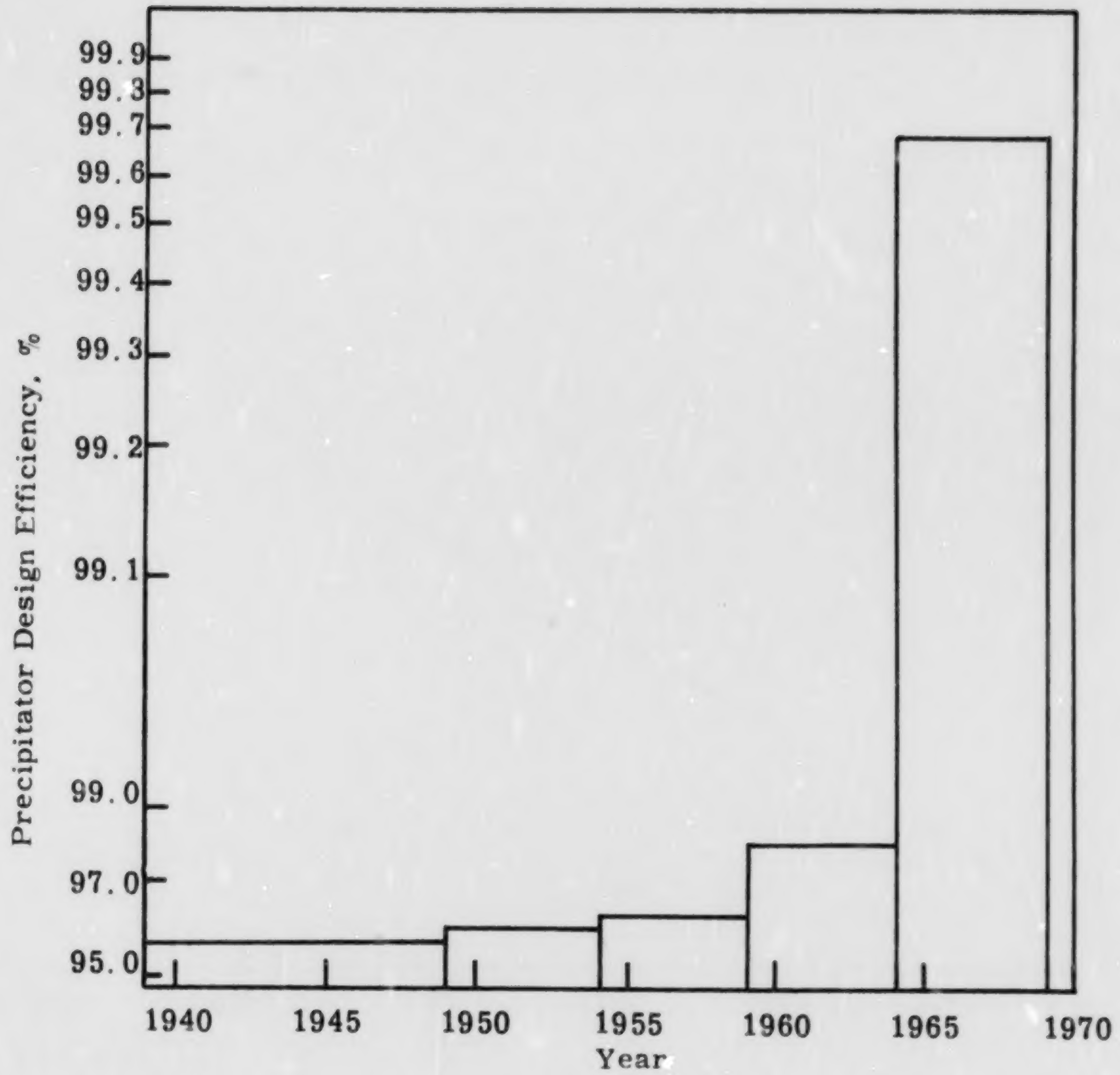

Figure 18.8. Design Efficiency Trend for Cement Plant Precipitators (1912-1969).

Table 18.5

Summary of Precipitator Design Efficiencies
(1949-1969)

	Wet Process		Dry Process		Combination Mills, Dryers, Vents	
Interval %	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm
>98.0	2	0.392	1	0.111	1	0.178
98.0-98.5						
98.5-99.0	2	0.311	1	0.254	1	0.049
99.0-99.5	1	0.129	2	0.138		
99.5+	8	1.520	1	0.185	1	0.014
Totals	13	2.352	5	0.688	3	0.241

Table 18.6

Summary of Precipitator Design Gas Velocities for Cement Industry Precipitators (1949-1969)

	Wet Process		Dry Process		Combination Mills, Dryers, Vents	
Interval (fps)	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm
3-4	5	0.880	3	0.323	1	0.014
4-5	6	1.080	1	0.111		
5-6	1	0.307			2	0.227
6-7			1	0.254		
7-8						
8-9	1	0.085				
Totals	13	2.352	5	0.688	3	0.241

Table 18.7

Summary of Precipitator Design Input Powers for Cement Industry Precipitators (1949-1969)

Interval (watts/10^3 acfm)	Wet Process		Dry Process		Combination Mills, Dryers, Vents	
	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm
50-100	4	0.713	2	0.365	1	0.049
100-150	4	0.847	1	0.078		
150-200	2	0.334			1	0.178
200-250	1	0.129	1	0.060		
250-300					1	0.014
300-350	2	0.329				
350-400						
Totals	13	2.352	4	0.503	3	0.241

Table 18.8

Cement Industry Summary of Statistics (1949-1969)
Precipitator Design Avg. Field Strength

	Wet Process		Dry Process		Combination Mills, Dryers, Vents	
Interval (kV/in.)	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm	Number of Installations	10^6 acfm
5-6	1	0.369				
6-7	3	0.757				
7-8	1	0.162				
8-9	1	0.370				
Totals	6	1.658				

Table 18.9

Summary of Performance Statistics on Cement Plant Electrostatic Precipitators

	Critical Parameter	Wet Process Kiln		Dry Process Kiln		Mill, Dryer, Vent Gases	
		Max	Min	Max	Min	Max	Min
1.	Gas Volume/Precipitator (acfm in thousands)	369	28	127	15	178	6.9
2.	Precipitator Efficiency (per cent)	99.94	92.8	99.98	98.79	99.91	96.90
3.	Gas Velocity in Precipitator (fps)	8.1	3.0	6.6	3.2	5.5	3.9
4.	Precipitator Inlet Dust Conc. (gr/scfd)	48.4	4.1	53.6	3.7	40.6	7.5
5.	Precipitator Input Power (watts/1000 acfm)	333	60	210	61	277	95
6.	Precipitator Avg. Field Strength (kV/in.)	8.8	5.2	-	-	-	-
7.	Gas Moisture (per cent by volume)	41.4	13.4	14.0	4.0	7.5	1.5
8.	Gas Temperature (°F)	650	353	700	500	160	93
9.	Precipitator Performance Ratio (R)	2.20	0.66	1.43	0.99	2.09	0.98

Design considerations. The design methodology for precipitators for cement plants follows the same practice as for other precipitator applications. The procedures outlined in Chapter 9 generally apply to cement plant precipitators, using design parameters that apply to the particular process.

The size precipitator required to achieve a given efficiency with a specified gas flow can be computed using the Deutsch-Anderson equation if the precipitation rate parameter is known. This parameter is governed principally by the nature of the dust and gas handled and, in the case of cement dust, is influenced principally by the dust particle size and resistivity. The particle size can vary with the operating parameters of the kiln and the level of alkali present in the rock. Resistivity is governed principally by moisture content and temperature.

Figure 18.9 is a plot of efficiency versus the collecting surface area to gas volume ratio. Design values are shown on the curve, and the shaded area represents the spread in design values for a group of 8 cement plant precipitators. Both wet and dry processes are represented by the data. From the limited data available, no difference is apparent in the design values used for the two processes. The precipitation rate parameter is the slope of the curve, and the data show values ranging from about 0.25 ft/sec (7.5 cm/sec) to around 0.45 ft/sec (13.5 cm/sec). Data from a group of 7 tests are also plotted on the same curve to show the variations from design values. Table 18.10 shows the design data for two cement installations. Table 18.11 contains design and performance precipitation rate parameters for cement plant electrostatic precipitators on wet process, dry process, and mill exhaust applications.

Power. The installed power supply capacity for cement industry applications varies over a rather wide range. There were eight installations for which installed power supply capacity is available. These range from about 250 to 1000 watts per thousand cfm. Figure 18.10 shows the range of installed capacity reported for the eight installations, together with delivered power for other installations.

Electrostatic precipitator problems. A problem associated with the use of electrostatic precipitators in collecting cement kiln effluents is the formation of crusty buildup on the internal numbers of the precipitator. This problem is reportedly due to the sulfur introduced either in the fuel, or feed to the kiln. The sulfur reportedly combines with the alkali or lime to form compounds that dissociate at kiln temperatures and recondense as solids on the internal surfaces of the precipitator if the temperature is

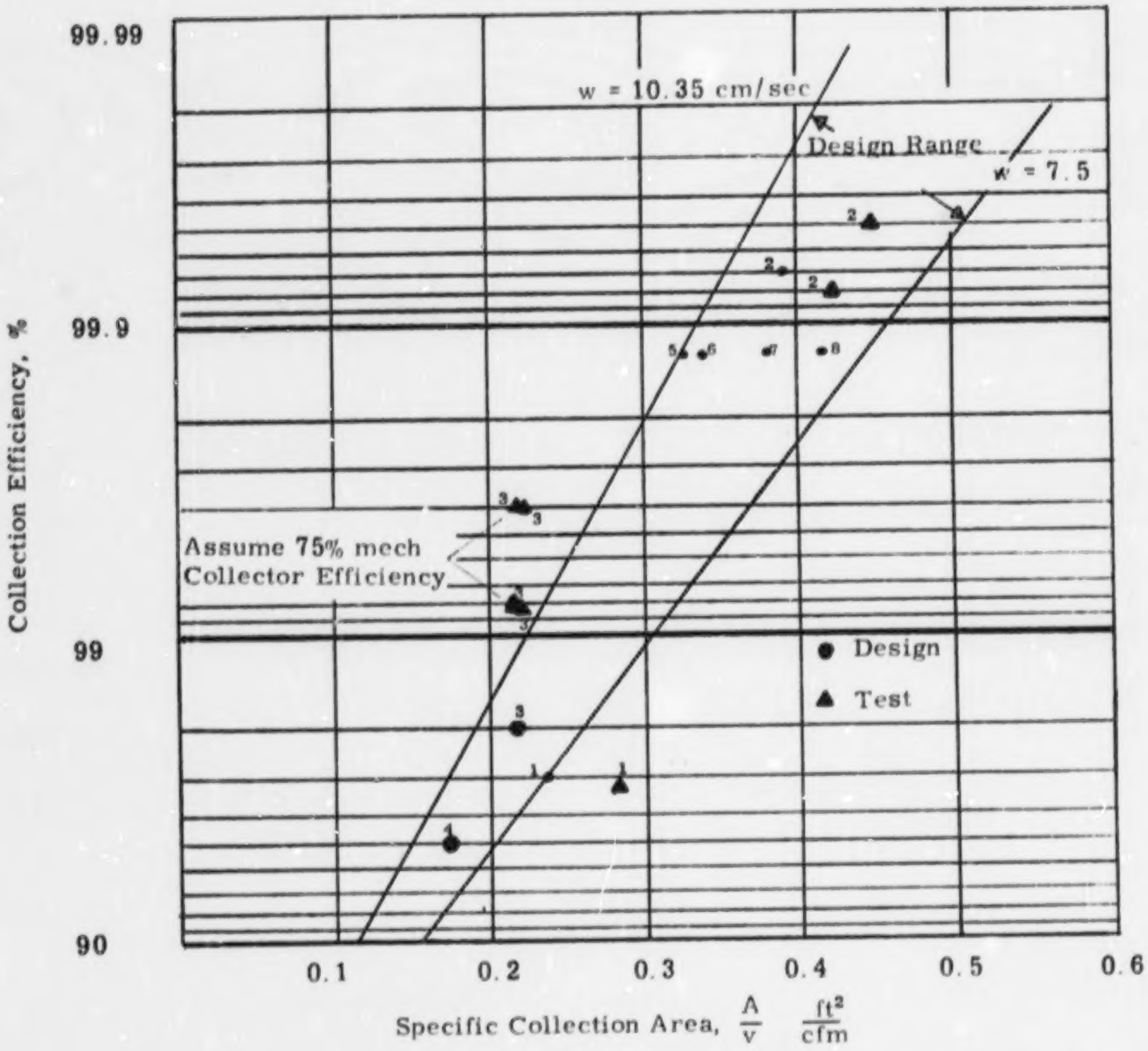

Figure 18.9. Relationship Between Collection Efficiency and Specific Collection Area for Cement Industries.

Table 18.10

Design Data for Two Current Installations

Process	Wet	Dry
Gas volume	275,000	122,000
Inlet Gas Temperature (°F)	500	600
Inlet Dust Loading (gr/ft^3)	14	16.8
Exit Dust Loading (gr/ft^3)	0.02	0.05
Efficiency (%)	99.86	99.6
No. of Gas Passages	62	29
Treatment Time (sec)	7.2	4.35
Collecting Plate Height (ft)	24	20
Collecting Plate Area (ft^2)	89,280	26,000
Length of Discharge Electrode (ft)	59,520	17,350
Ratio of Plate Area to Gas Volume (ft^2/1000 cfm)	325	218
Ratio of Wire Length to Gas Flow (ft/1000 cfm)	216	142
No. of Rectifiers	4 (70kV-750mA)	2
No. of Sections	8	
Current Density (mA/1000 ft^2)	34	32
No. of Plate Rappers	36 (24 ft-lb)	36
No. of Discharge Electrode Rappers (vibrators)	16	16
Precipitation Rate Parameter w (ft/sec)	0.34	0.3
Power Density (watts/thousand cfm)	425	225

Table 18.11

Design and Performance Precipitation Rate Parameters for Cement Plant Electrostatic Precipitators

	Design w (ft/sec)			Performance w (ft/sec)		
	Max	Min	Avg.	Max	Min	Avg.
Wet Process (6 precipitators)	0.41	0.29	0.35	0.47	0.27	0.38
Dry Process (4 precipitators)	0.34	0.23	0.31	0.43	0.25	0.35
Mill Exhaust (2 precipitators)	0.25	0.25	0.25	0.51	0.38	0.45

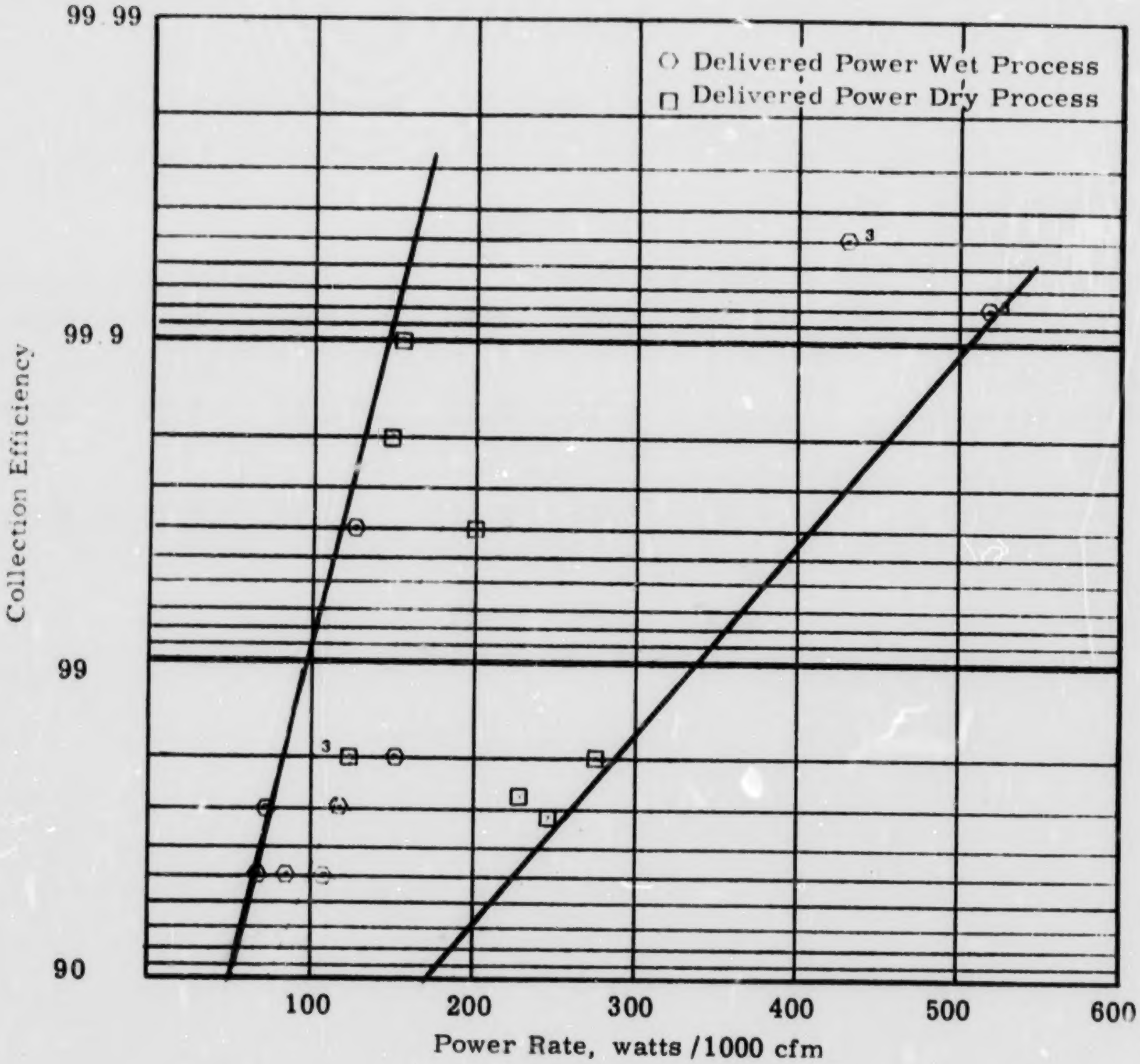

Figure 18.10. Efficiency Vs. Power Rate for Cement Plants.

reduced below a certain critical value. Depending on the temperature, these deposits occur either as heavy dense formations or solid plate-like deposits that are difficult to remove. This sulfate buildup can be prevented by cooling the gases below the condensation temperature before entering the precipitator (around 550°F).

The nature of the kiln gases is such that corrosion is a problem if moisture condensation occurs on the steel surfaces of the precipitator. This means that gas temperatures within the precipitator should be considerably above the dew point, and the shell should be insulated to minimize corrosion problems. In some European installations, which operate at such low temperatures that condensation does occur, aluminum alloy has been successfully used for the internal parts of the precipitators.

Another problem associated with the use of electrostatic precipitators with cement kilns, or any other effluent from a combustion process, is the danger of combustion in the precipitator. If a significant amount of combustible material remains in the kiln effluent, and if sufficient fresh air enters with this gas, either through leaks or for cooling purposes, the sparks in the precipitator or from other sources can ignite the combustible mixture. The heat evolved from this combustion will raise the temperature sufficiently to melt or deform the internal parts of the precipitator so badly that it cannot be used until these parts are renewed. Such fires are normally prevented by use of controls operated by instruments that monitor the concentration of excess oxygen, concentration of combustibles, and the temperature of the exit gases from the kiln ahead of the precipitators. A dangerous increase in combustible content or temperature sets off alarms and shuts off the fuel to the kiln burners, and in at least one installation, closes sliding doors on the precipitator inlet and outlet.

The problem of combustible mixtures is not limited to electrostatic precipitators, but since the sparking normally associated with precipitators is a potential source of ignition, the danger of fires is perhaps amplified.

Rapping of the dust from electrostatic precipitators operating on cement dust is difficult due to the fineness of the dust and the moisture content. Dust removal in this instance usually requires more severe and frequent rapping than for other applications.

Economics. Precipitator FOB costs for cement plants vary from around \$0.50 to \$1.00 per cfm, whereas erected costs range from around \$1.00 to over \$3.00 per cfm. Figure 18.11 shows the range of FOB costs as a function of gas volume handled for a group of 7 installations. The spread in erected cost data is too great for the limited number of installations to show statistically meaningful trends. The range for FOB costs is from \$0.60 to \$0.72 per cfm. Erected costs for the same installations range from \$1.30 to \$1.63 per cfm. Another price comparison is the cost per square foot of collection area. This range is from \$1.60 to \$2.10 for FOB costs, and from \$3.50 to \$6.00 for erected costs.

Maintenance cost data for precipitators installed on cement kilns are given in Table 18.12. The data for the four plants reflect the variation in the maintenance schedule and the estimated man hours required to perform maintenance services.

Power costs for the precipitator are estimated on the basis of the installed capacity, power costs of \$0.0075/kWh, and 8000 hours operation per year. Fan costs are based on the gas volume and a pressure drop of 0.5 in. water. Of the four plants, none indicated labor allocation for precipitator operation.

18.4 ELECTROSTATIC PRECIPITATORS FOR GYPSUM INDUSTRY

Gypsum is a hydrated calcium sulfate ($CaSO_4 \cdot 2H_2O$) which occurs in natural deposits in many areas throughout the world. When heated to a temperature in the range of 250-300°F, the gypsum loses 1-1/2 moles of water to produce a product called "first-settle plaster" or plaster of Paris; further heating to a temperature of around 375°F results in the loss of the remaining 1/2 mole of water. The anhydrous calcium sulfate is called "second-settle plaster," and is the product used as a building material in the form of plaster and plaster board.

Removal of the water of hydration in the production of plaster is a process termed "cooking," which takes place in either a kettle or in a rotary kiln. The gypsum kettle is a large vertical cylinder, about 8-10 ft in diameter and 8-10 ft in depth, contained in an insulated jacket. The space between the kettle and jacket serves as a combustion chamber, where oil or gas is burned to provide heat for the process. The kettle is equipped with a motor-driven agitator to stir the charge during cooking.

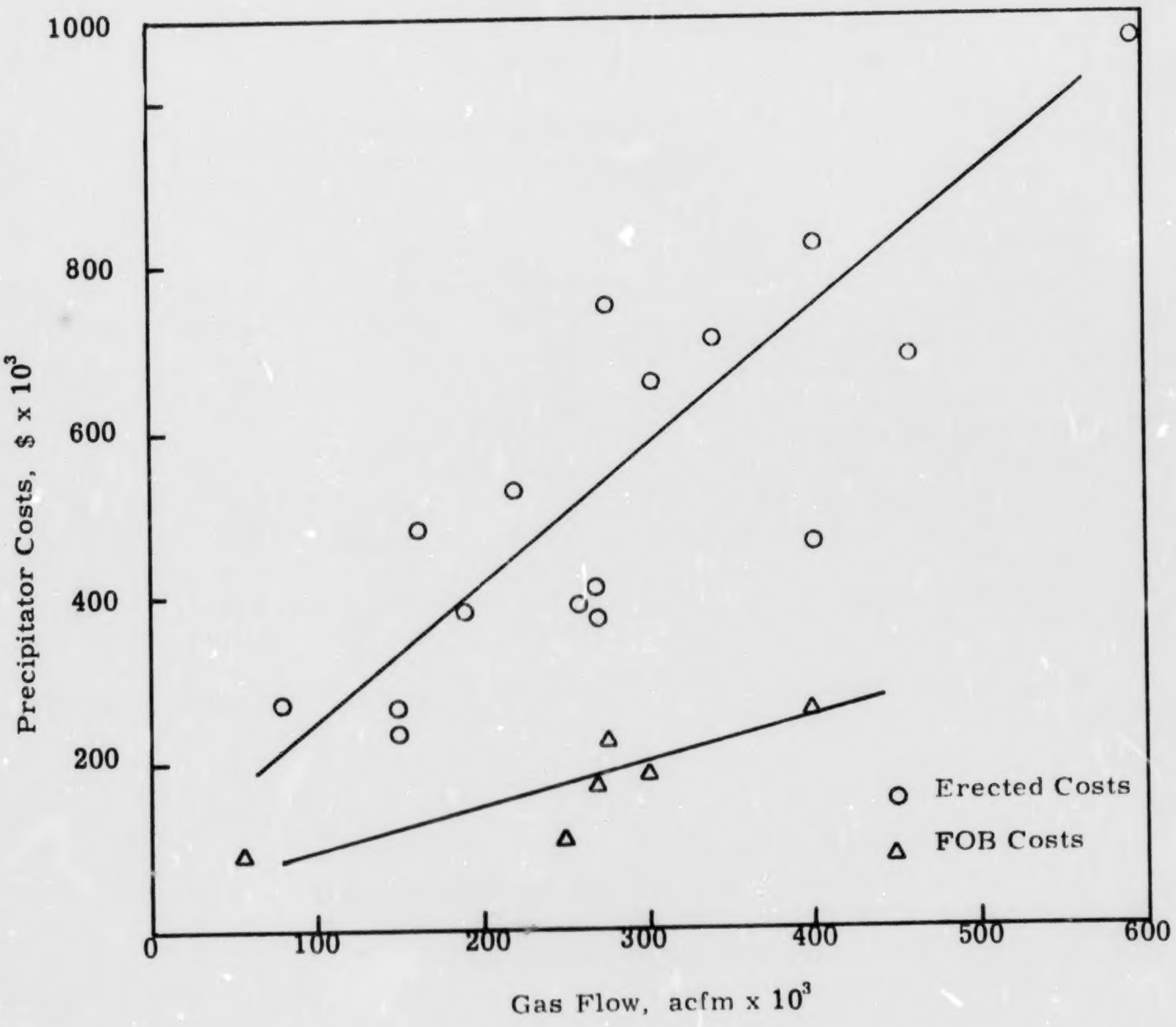

Figure 18.11. FOB and Erected Costs for Precipitators in the Cement Industry for Collection Efficiencies from 98-99.9%.

Table 18. 12

Maintenance and Operating Costs for Precipitators Installed in Cement Plants

Plant No.	1	2	3	4
Capacity, BBl/day	8, 000	2, 900	2, 700	8, 500
Gas Flow, cfm	275, 000	120, 000	58, 000	250, 000
No. of Maintenance Periods/yr.	3-4	6-10	6-8	2-3
Man hours/period	16		40	60-100
Avg. Maintenance, Man hours/yr.	80		275	200
Operating Man hours	--	--	--	--
Maintenance Labor Costs [1]	480		1, 650	1, 200
Precipitator Power, kW	50	30	13	60
Precipitator Power Costs [2]	3, 000	1, 800	780	3, 600
Fan, kW [3]	17. 2	7. 5	3. 6	15. 5
Fan Power Costs[4]	1, 035	450	216	935
Capital Costs [4]	64, 000	34, 500	20, 500	59, 000
Total Annual Costs	68, 515	36, 750 [5]	23, 146	64, 735

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[1] Based on $6. 00 per hour

[2] Based on Electricity Costs at $0. 005 per kWh at 8, 000 hours/yr

[3] Based on 1/2" W. G. pressure drop for precipitator

[4] Based on 12. 8% of the Erected Costs per year

[5] Maintenance Labor Costs not shown

The charge for the kettle is raw gypsum rock. This rock first passes through a dryer at a temperature of 230°F where excess moisture is removed. The rock is then crushed, milled, and ground to a coarse powder prior to feeding it into the kettle. The pulverized gypsum from the grinding mill is pneumatically fed to a cyclone separator where the kettle feed is separated from the transport air and fed into a storage bin ready for charging into the kettle.

The cooking process requires about 2-5 hours for completion. When completed, the charge is transported through a chute to a hot pit located below the kettle. Figure 18.12 is a block diagram showing the steps in the process.

Gypsum is also produced in a continuous kiln called a "Holo-Flight" calciner. The kiln is equipped with a screw conveyor for transport of the material. The shaft, the conveyor, and the screw flights are hollow, and hot oil is pumped through them to provide the heat for calcining.

Particulate emissions and characteristics. Particulate emissions from processing of gypsum cover a wide range, as shown in Table 18.13. The factors influencing dust emission are the rate of calcination, the size of the gypsum, and the extent of the stirring or agitation.

The size of the dust particles from the gypsum processing plant is large compared with that of cement or other higher temperature processes, since the majority of the dust is produced as a result of mechanical abrasion and gas release during calcination. Figure 18.13 shows typical particle size distribution ranges.

The moisture content of the gases from the calcining process is high, since water of hydration is released from the gypsum rock. Each ton of gypsum calcined releases approximately 314 lb of water.

The exiting wet kettle gas is usually mixed with some flue gas to increase its temperature. Before entering the precipitator, ventilating air from several sources in the plant is often combined with the kettle gas in a mixer. The temperature of the final gas mix entering the precipitator ranges from 300°F to 400°F. The moisture content ranges from 30% to 35% water vapor by volume. Table 18.14 shows the range of moisture contents for the various processes. The dust concentration entering the precipitator varies widely depending on the cycle phase of the batch; the variation is from essentially 0 to 50 grains of dust per standard cubic foot of dry gas.

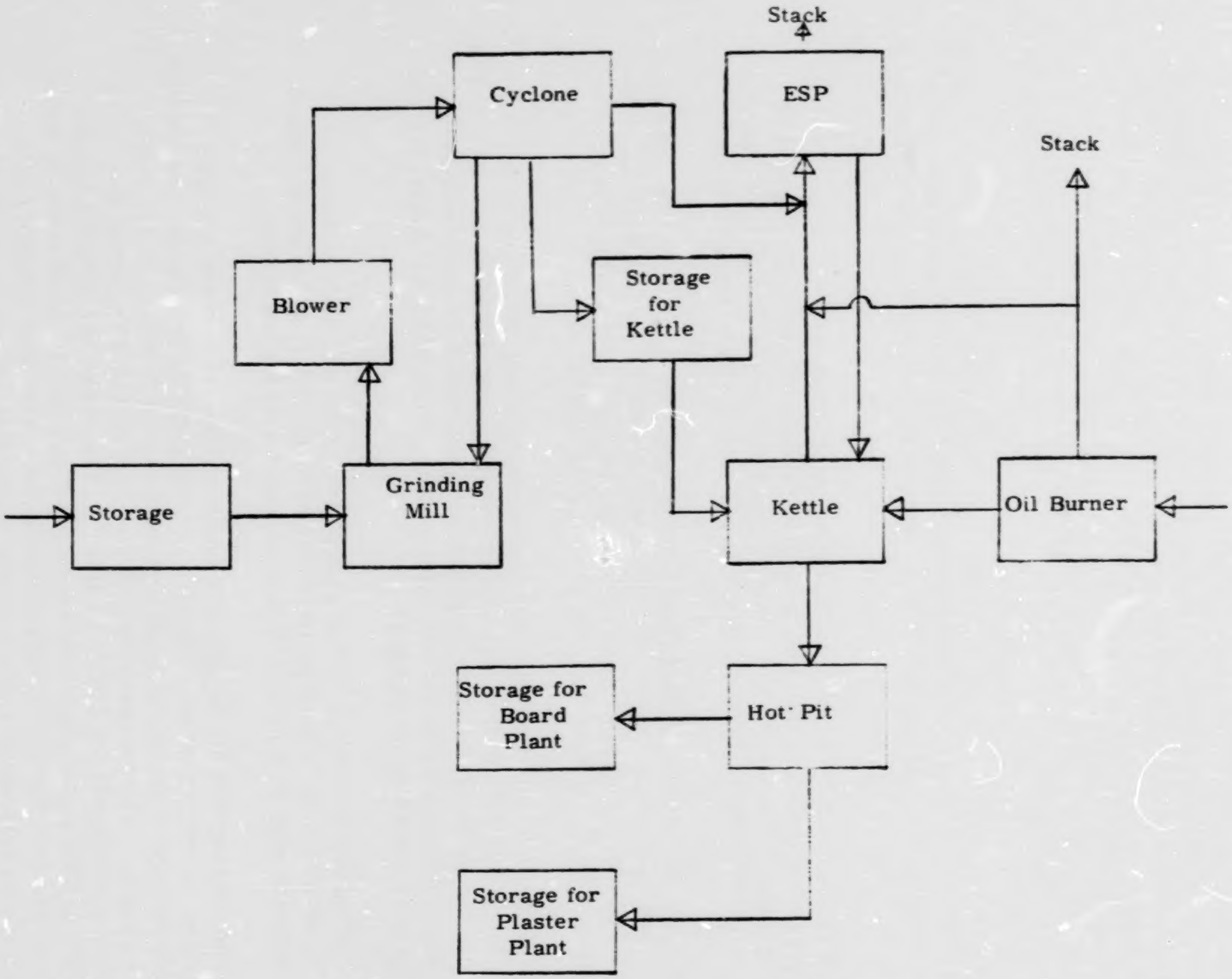
Stack
ESP
Cyclone
Stack
Blower
Storage
for
Kettle
Storage
Grinding
Mill
Kettle
Oil Burner
Storage for
Board
Plant
Hot Pit
Storage for
Plaster
Plant

Figure 18.12. Schematic Diagram of a Gypsum Plant with Electrostatic Precipitator.

Table 18. 13

Precipitator Inlet Dust Loadings for Gypsum Industry Precipitators (1935-1969)

Interval Pptr. Inlet Dust Loading (gr/scfd)	Kettles		Rock Dryers		Calciner (Rotary)		Calciner (Holoflite)		Combination Kettles, Dryers Mills	
	Number	10^3 acfm	Number	10^3 acfm	Number	10^3 acfm	Number	10^3 acfm	Number	10^3 acfm
<20.0	3	48.52	4	68.38					10	314.18
20-30	1	5.9							2	42.9
30-40	2	20.85			4	199.65	1	11.3	2	64.2
40-50	1	9.2			1	31.80			3	107.5
50-60							2	29.65		
60-70			3	86.9			1	8.7		23.0
>70.0			3	72.55						
Totals	7	84.47	10	227.83	5	231.45	4	49.65	18	551.78

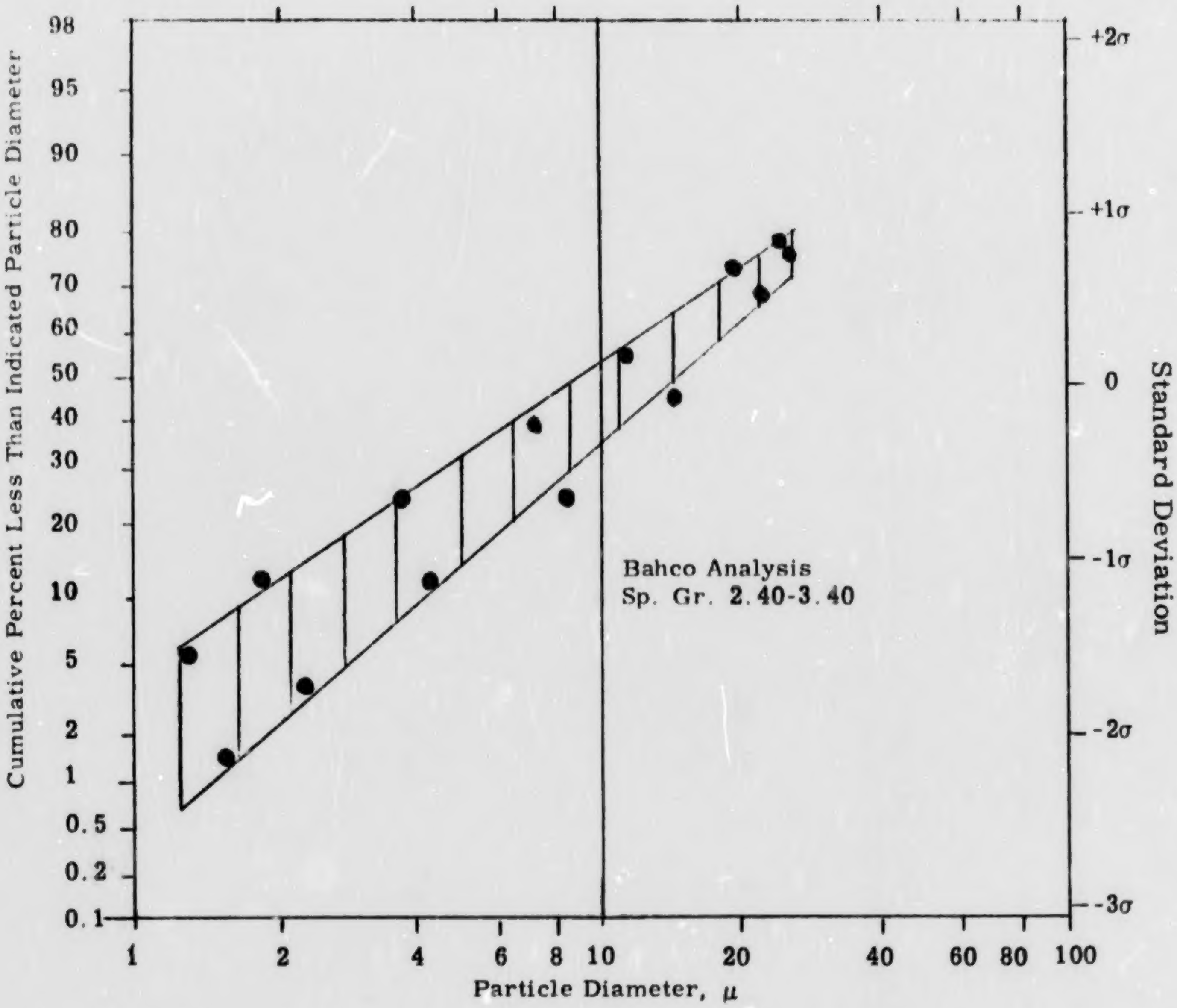

Figure 18.13. Typical Particle Size Distribution of Gypsum Dust.

Table 18.14

Range of Gas Moisture Contents to Precipitators in Gypsum Plants

(1935-1969)

Interval	Kettles		Rock Dryers		Calciner (Rotary)		Calciner (Holoflite)		Combination Kettles, Dryers Mills	
(% by Vol)	Number	10^3 acfm	Number	10^3 acfm	Number	10^3 acfm	Number	10^3 acfm	Number	10^3 acfm
0-10			3	73.5					3	66.9
10-20	1	7.27	6	129.73	1	31.8	1	16.45	10	327.85
20-30	5	69.07			2	84.1			3	77.78
30-40	1	7.55			2	115.55	2	29.8		
40-50	2	17.35					1	8.7		
50-60							2	18.88		
Totals	9	106.67	9	203.23	5	231.45	6	73.83	16	472.53

Dry gypsum or plaster dust has a resistivity which could potentially be high enough to present a collection problem. However, since the dry vent gases are usually combined with the moisture-laden gases from the calcining process before they enter the precipitator, the problem of high resistivity is usually avoided, and electrostatic precipitation of gypsum dust has been a satisfactory control application. At times, however, in addition to the wet kettle gases, the dryer mixed gases must be humidified by the addition of water sprays in the flue ahead of the precipitator. Figure 18.14 is a plot of laboratory and in-situ resistivity versus temperature for two moisture contents. Table 18.15 shows the range of inlet gas temperatures which, along with moisture, determines dust resistivity.

Precipitator design and performance. Table 18.16 shows the number of electrostatic precipitators installed in gypsum plants during the period 1935-1969. Table 18.17 shows the application of these precipitators, along with the total gas volume handled, and the average gas volume per precipitator. Figure 18.15 shows the gas volume of precipitators installed at 5-year intervals between 1935-1970, and the cumulative gas volumes over this period. The proportion of the capacity installed on each type of equipment is also shown.

Two types of electrostatic precipitators are in general use for dust control in gypsum plants. An earlier model that is still being used is the vertical-flow, single-stage precipitator, which employs hollow-core collecting electrodes in conjunction with horizontal rod-type discharge electrodes. The newer installations utilize the standard single-stage, horizontal-flow, duct design using shielded flat-plate collecting electrodes.

Table 18.18 lists some of the design and performance factors for precipitators collecting dust from gypsum kettles, calciners, and combination kettle, calciner, dryer, and grinding mills.

The design parameters for gypsum installations presented in Figure 18.16 are based on data from two installations. These installations were equipped with mechanical collectors before the electrostatic precipitator. The precipitation rate parameters presented were computed from the Deutsch-Anderson equation based on an assumed collection efficiency of 75% for the mechanical collector.

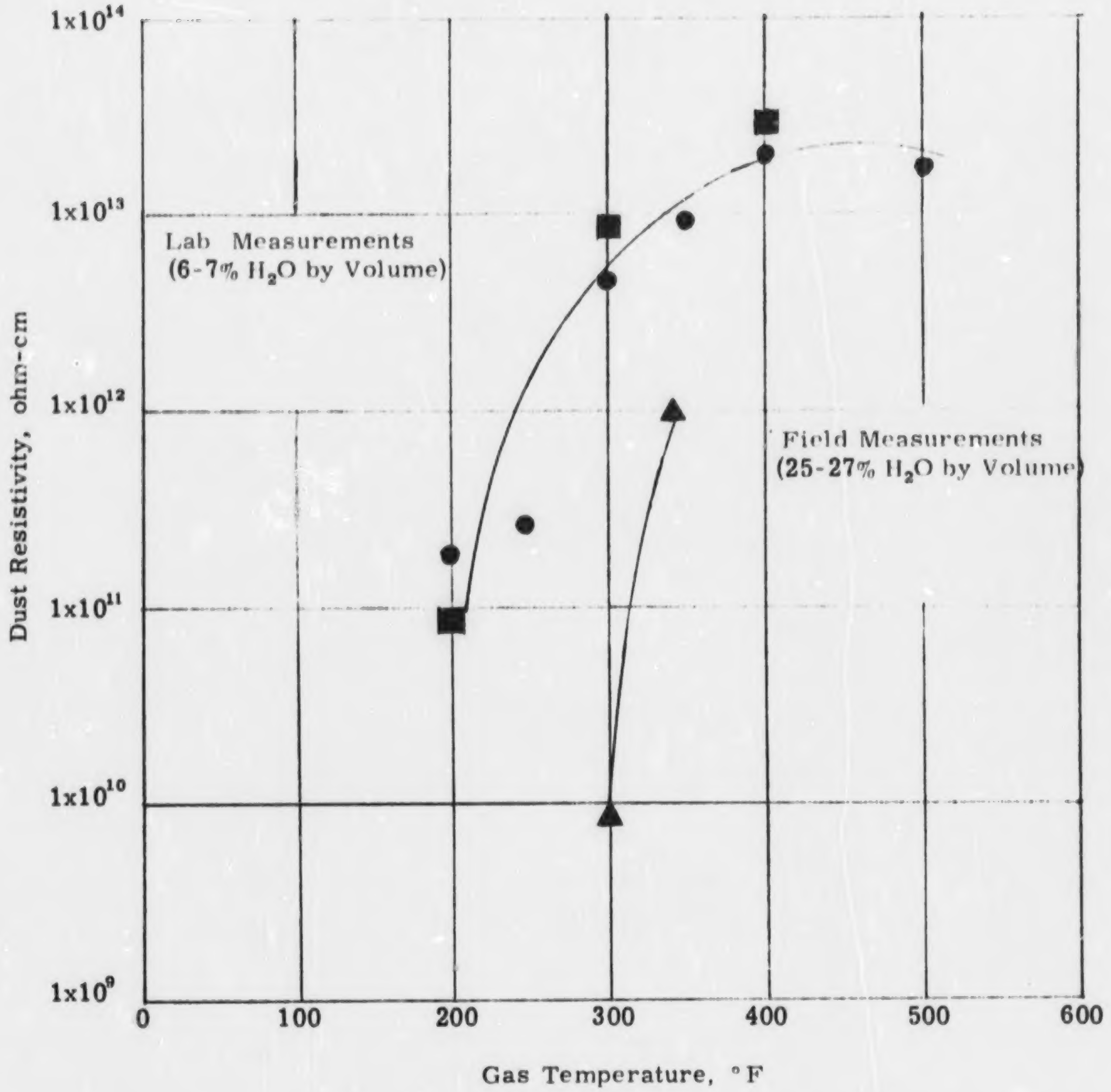

Figure 18. 14. Typical Laboratory and Field Resistivities of Gypsum Dust.

Table 18.15

Summary of Precipitator Gas Temperatures for Gypsum Industry Precipitators (1935-1969)

Interval	Kettles		Rock Dryers		Calciner (Rotary)		Calciner (Holoflite)		Combination Kettles, Dryers Mills	
(°F)	Number	10^3 acfm	Number	10^3 acfm	Number	10^3 acfm	Number	10^3 acfm	Number	10^3 acfm
100-150			2	70.9					2	74.4
150-200			3	72.55					8	276.93
200-250	1	12.7	1	25.5			2	33.05	4	119.8
250-300	2	21.5					4	41.0	1	29.38
300-350	2	23.8					1	7.58		
350-400										
Totals	5	72.55	6	168.95			7	81.63	15	500.51

Table 18.16

Electrostatic Precipitator Installations in Gypsum Plants
(Period 1935-1969)

Pptr. Contract Year (s)	Five (5) Year Period Indicated No. of Install.	No. of Pptrs.	Total Gas Volume (10^3 acfm)	Average Volume/Year During Period (10^3 acfm)	Weighted Design Efficiency on acfm Basis (%)
1935-1939	4	5	100	20	98.5
1940-1944	3	3	56	11.2	98.6
1945-1949	3	3	77.5	15.5	96.5
1950-1954	18	25	545.5	109	97.5
1955-1959	20	21	607.6	121.5	97.7
1960-1964	10	11	351.6	70.3	99.15
1965-1969	1	2	80	16.0	99.0
Grand Totals	59	70	1818.2		

NOTES: (1) The statistics in this table include precipitators on rock dryers, kettles, rotary calciners, holoflite calciners, and combinations of calciners and kettles with dryer and grinding mill vent gases.

(2) Table 18.15 breaks the data down into the various categories.

Table 18.17

Electrostatic Precipitator Installations in Gypsum Plants
Listed by Area of Application
Period 1935-1969

Application	Number of Installations	Number of Precipitators	Total Volume (10^3 acfm)	Volume/Precipitator (10^3 acfm)
Kettles	11 (30)	19	164.2	8.6
Rotary Calciners	6 (8)	6	285	47.5
Holoflite Calciners	8 (8)	8	211.6	26.5
Rock Dryers	10 (10)	10	252	25.2
Combinations of Kettles, Dryer and/or Mills	22	25	902	36.1

NOTE: Numbers in parenthesis are individual pieces of indicated equipment.

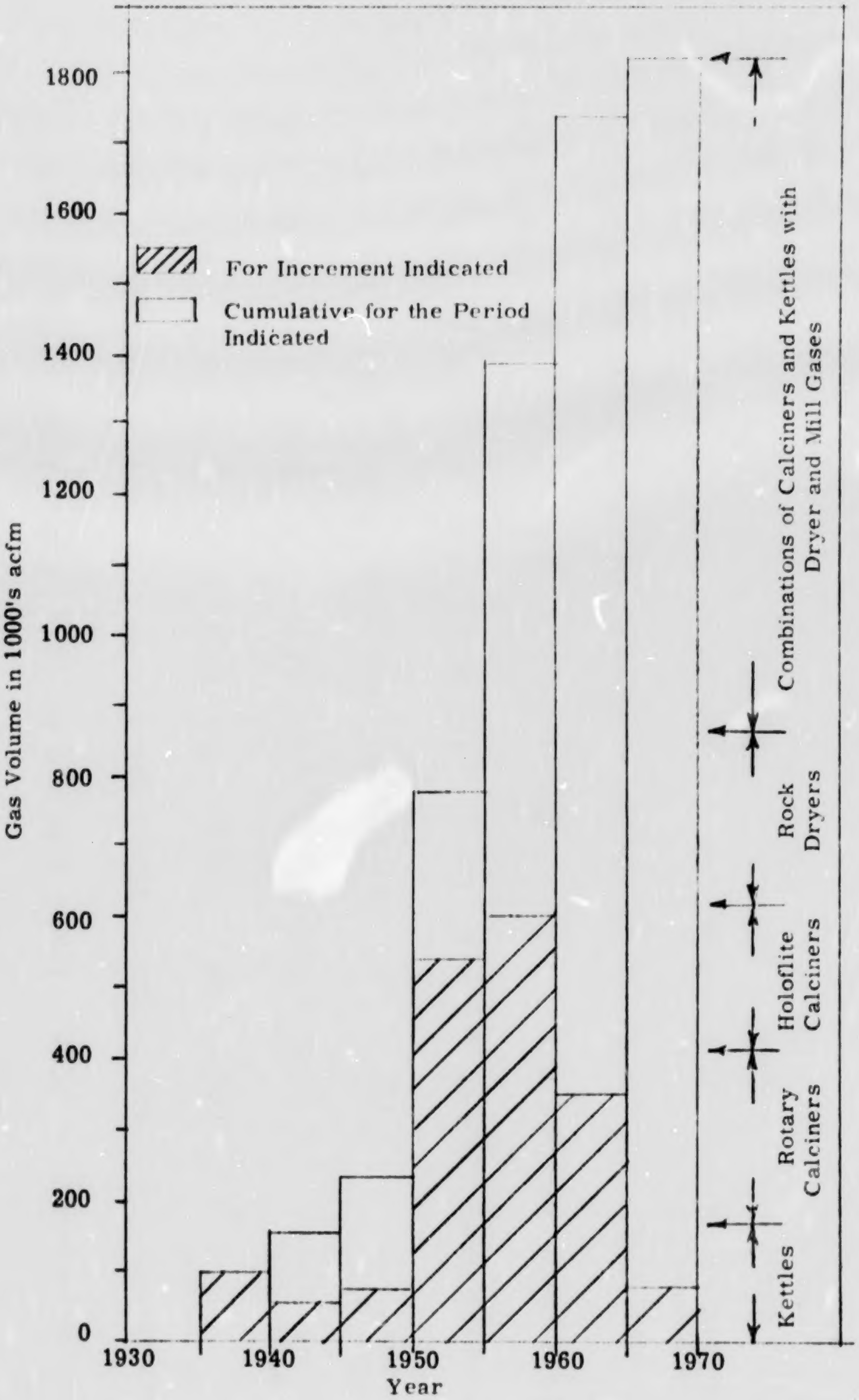

Figure 18.15. Electrostatic Precipitators in the Gypsum Industry (Installed Capacity) Data from Tables 18.14 and 18.15.

Table 18.18

Summary of Performance Data on Gypsum Plant Electrostatic Precipitators

Critical Parameter	Kettles		Rock Dryers		Calciner (Rotary)		Calciner (Holoflite)		Combination Kettle, Dryer, Mill	
	Max	Min	Max	Min	Max	Min	Max	Min	Max	Min
1. Gas Volume/Precipitator (acfm in thousands)	13.6	2.9	43.0	10.8	82.5	29.8	16.6	7.6	60.2	9.0
2. Precipitator Efficiency (per cent)	99.90	94.38	99.85	97.20	99.82	99.14	99.90	98.20	99.96	93.76
3. Gas Velocity in Precipitator (fps)	5.5	1.5	7.5	3.1	7.9	4.2	2.8	1.4	7.4	3.0
4. Precipitator Dust Concentration (gr/scfd)	48.0	4.8	156	4.6	46.3	32.9	62.6	30.8	63.9	7.7
5. Precipitator Input Power (watts/1000 acfm)	828	124	282	69	174	62	983	576	390	65
6. Precipitator Avg. Field Strength (kV/in.)	12.8	8.8	15.0	9.0	13.0	8.1	15.5	9.8	13.4	8.0
7. Gas Moisture (per cent by volume)	47.6	19.0	17.4	3.6	35.2	14.2	54.0	17.6	22.2	6.7
8. Gas Temperature (°F)	342	220	240	125	-	-	339	200	250	140
9. Precipitator Performance Ratio (R)	2.39	0.74	2.06	0.89	1.45	1.03	1.39	0.47	2.39	0.40

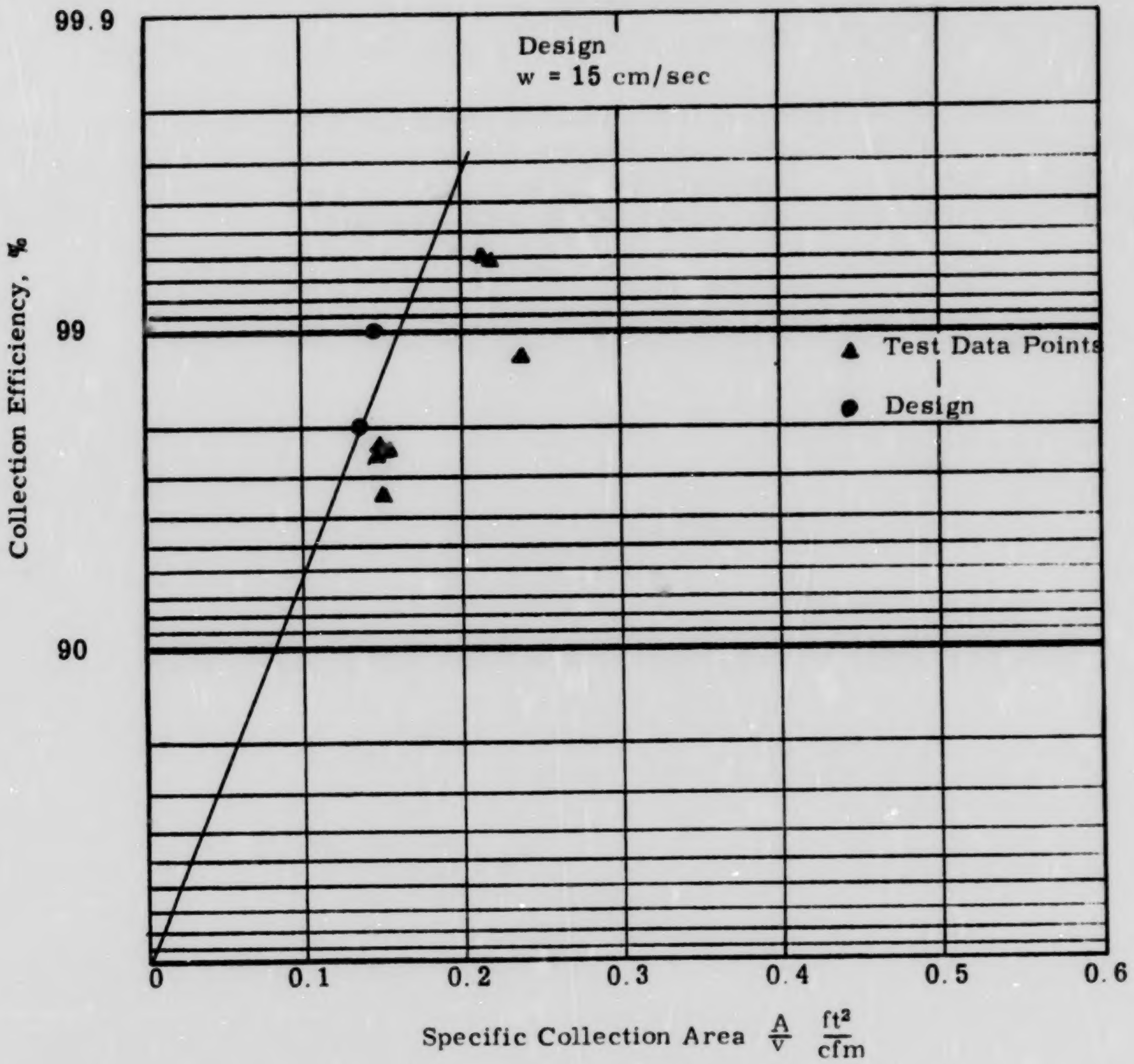

Figure 18.16. Collection Efficiency Vs. Specific Collecting Surface for Gypsum Rock Dryer Plant Preceded by Mechanical Collector. Efficiency of Mechanical Collector Assumed to be 75%.

Economics. Table 18.19 shows erected and FOB costs for precipitators installed in gypsum plants over the period 1959-1967. The costs per acfm range from $0.89 to $1.03 for 99.0% collection efficiency, and from $1.43 to $1.53 for 99.5%. Erected costs vary from $1.38 to $3.62 per acfm.

Table 18.20 lists the maintenance costs for two precipitators installed in gypsum plants. The fan costs were based on a pressure drop of 1/2" W.G. in the precipitator. The plant was assumed to operate 8000 hours per year with electricity costs estimated at $0.0075 per kWh.

Table 18.19

Gypsum Industry Economics

Year	Type Cost	Design Volume (1000's acfm)	Design Efficiency (%)	Total Cost (10^3 $)	Cost/Unit Volume ($/acfm)
1959	FOB	40	99.0	39.5	0.99
	Erected	40	99.0	55.2	1.38
1960	FOB	40	99.0	38.7	0.97
1961	Erected	40	99.7	90.1	2.25
1962	FOB	40	99.0	35.7	0.89
1964	FOB	36	99.5	55.0	1.53
		36	99.5	51.3	1.43
		36	99.5	56.3	1.57
		40	99.0	36.0	0.90
1965	Erected	27	99.0	97.8	3.62
1967	FOB	40	99.0	41.0	1.03

Table 18.20

Maintenance Data for Precipitators Installed on Gypsum Calciners

Prod. Rate tons/hr.	17-17	60-90
Gas Flow, acfm	16,600	35,000
Precipitator Power, kW	12.2	13.5
No. of Maintenance Periods/year	3-4	3-4
Man hours/year	70-90	70-90
Maintenance Labor Costs[1]	480	480
Fan Power, kW[2]	1.04	2.19
Power Costs - Fan[3], $/yr.	62.25	131.25
Power Costs - Precipitator, $/yr.	730	810
Capital Costs,[4] $/yr.	8,600	9,350
Total Annual Costs	9,872.25	10,771.25

[1] Based on $6.00 per hour

[2] Based on 1/2" W.G. pressure drop

[3] Based on Power Costs at $0.0075 per kWh

[4] Based on 12.8% of total Erected Costs

CHAPTER 18
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CHAPTER 19
ELECTROSTATIC PRECIPITATORS IN THE CHEMICAL INDUSTRY

19.1 MAJOR FIELDS OF APPLICATION

The major applications of electrostatic precipitators in the organic chemical industry are in gas cleaning operations associated with the manufacture of sulfuric acid, phosphorus, and phosphoric acid. Historically, electrostatic precipitators have also been used in the production of carbon black, although this application has largely been replaced by other dust control methods.

In the production of sulfuric acid, electrostatic precipitators are used in cleaning of gases from ore roasters and in the collecting of acid mists from the effluent gas from both ore roasters and sulfur burning plants. In the manufacture of elemental phosphorus, precipitators are used to clean the gas prior to condensation of the phosphorus. Precipitators are used for acid mist collection in the manufacture of phosphoric acid. Use of precipitators in the manufacture of carbon black is primarily for agglomeration of the fines prior to collection in mechanical cyclones and fabric filters.

Sulfuric acid constitutes one of the largest volume industrial chemicals produced in the United States. In 1967, the production of sulfuric acid amounted to over 27 million tons. Its chief uses are in the production of fertilizer, manufacture of chemicals, oil refining, pigment production, iron and steel processing, synthetic fiber production, and in ferrous metallurgical operations.

Virtually all of the phosphatic fertilizer and phosphorus used in the U. S. come from phosphate rock in which phosphorus is present as fluorapitite $[(CaF_2 \cdot 3Ca_3(PO_4)_2]$. In 1964, there were over 66 million tons of phosphate rock mined in the U. S., with Florida producing nearly 90% of the total.

Carbon black is produced by the incomplete combustion of hydrocarbon in oil and gases. It is used as a component of rubber for automobile tires and as pigment in ink and paint products.

19.2 SULFURIC ACID PRODUCTION PROCESSES

Essentially all sulfuric acid is made by catalytic oxidation of sulfur dioxide to sulfur trioxide with subsequent absorption in a recirculating sulfuric acid solution to make acid of the desired concentration. Sulfuric acid is produced by two processes which are quite different in principle.

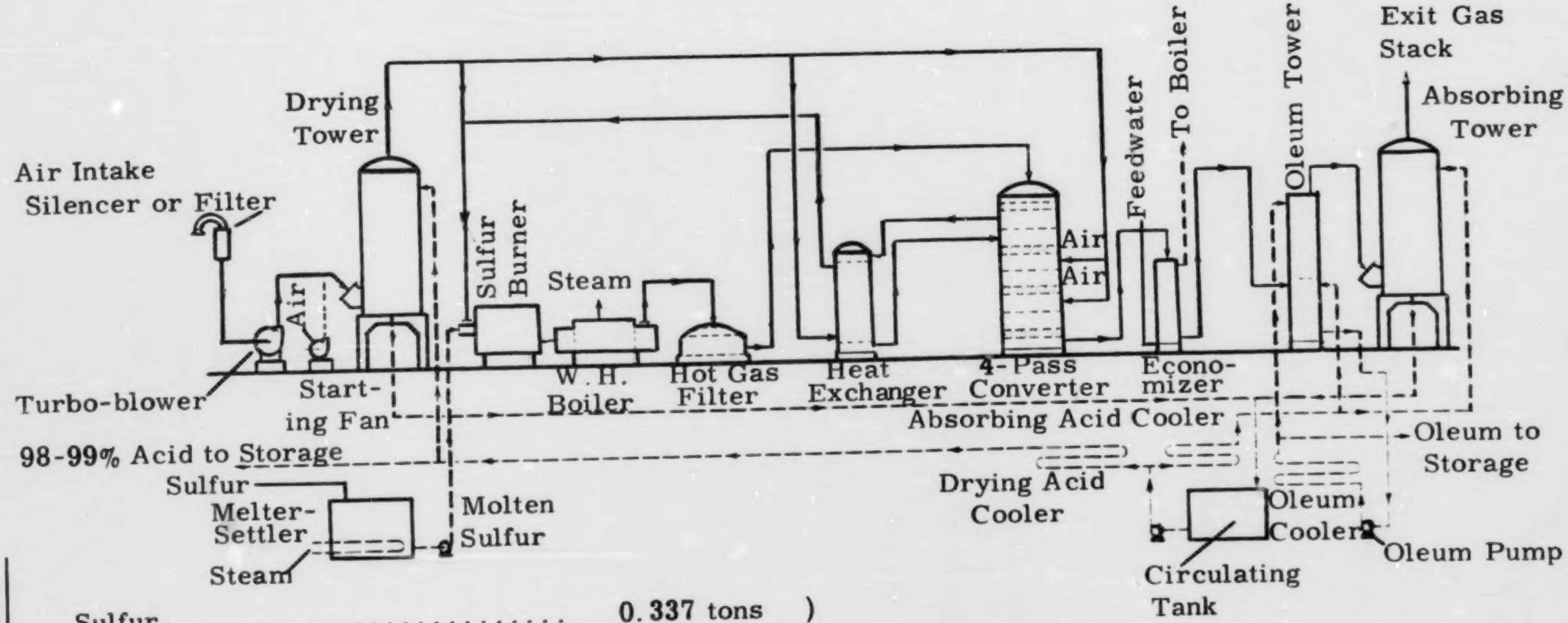
Chamber process. The oldest of the two methods for producing sulfuric acid is the chamber process. In this process, gaseous oxides of nitrogen serve as the catalyst, and oxidation takes place in lead-lined chambers. The chamber process produces sulfuric acid with a concentration of around 78%, which is sufficient for most needs of the fertilizer industry, such as the production of superphosphate. However, it is sometimes advantageous to produce a more concentrated acid by other processes, and dilute it as required.

In 1965, there were 60 chamber plants producing sulfuric acid in this country, mostly associated with fertilizer manufacture. However, these plants accounted for less than 5% of the total U. S. production. Because of the low concentration of the product acid, the high costs of operation, and the high emissions of nitrogen oxides, use of the chamber process is diminishing, and no new chamber process plants have been built since 1956.

Contact process. In the contact process, sulfur dioxide is catalytically oxidized to sulfur trioxide by atmospheric oxygen. Oxidation takes place in the converter, which is usually a cylindrical brick-lined shell containing trays of vanadium pentoxide catalysts. Oxidation of the sulfur dioxide takes place on the surface of the catalyst at atmospheric pressure, and at a temperature of approximately 1400° F.

The sulfur bearing raw materials for the contact sulfuric acid process can come from a variety of sources. Most of the sulfuric acid produced in the U. S. is made from elemental sulfur. The remaining new acid is made from pyrites; zinc, copper, and lead ores; smelter gas; hydrogen sulfide; and crude sulfur.

Burning of elemental sulfur. The combustion of sulfur for the contact process takes place in a brick-lined combustion chamber into which is fed either atomized molten sulfur or pulverized sulfur and combustion air. A flow chart for this system is shown in Figure 19. 1.


Air Intake
Silencer or Filter
Drying Tower
Air
Turbo-blower
Start-ing Fan
98-99% Acid to Storage
Sulfur
Melter-Settler
Steam
Molten Sulfur
Sulfur Burner
Steam
W. H. Boiler
Hot Gas Filter
Heat Exchanger
4-Pass Converter
Air
Air
Feedwater
To Boiler
Econo-mizer
Absorbing Acid Cooler
Drying Acid Cooler
Circulating Tank
Oleum Tower
Oleum Cooler
Oleum Pump
Oleum to Storage
Exit Gas Stack
Absorbing Tower

Sulfur	0.337 tons	Per ton of 100% acid in plant of 200 tons daily capacity
Water	4400 gal	
Electricity	10 kWh	
Direct operating labor	0.16 man-hr	
Steam credit	2000 lb	
Repairs (labor and materials)	60¢	

Figure 19.1. Typical Flowchart for Sulfur-Burning Contact Plant (Reference 4).

Because of the necessity for maintaining a low moisture content in the gases to the converter, the combustion air is dried by passing it through a tower where 98% sulfuric acid is used as a drying agent.

The combustion of sulfur is exothermic, and the gas leaving the sulfur burner is at a temperature in excess of about 1400°F. The gas which normally contains 8-10% SO_2 is cooled to around 800°F in the waste heat boiler before entering the converter.

Roasting of sulfide ore. When sulfide ores are the source of sulfur, the plant is usually more complex and expensive than when elemental sulfur is processed. The cost may be as much as three times that of the sulfur burning plant, and in addition, have lower yields. However, when the price of sulfide ore or smelter gas is low compared to that of elemental sulfur, the process may produce sulfuric acid at a lower cost.

In the United States, metallurgical plants account for approximately 15% of the total sulfuric acid production. In Europe, where elemental sulfur is not as readily available, a larger percentage of the sulfuric acid is from ore roasting plants, smelters, and other sulfur sources.

When smelter gases are utilized as a raw material, the sulfur dioxide content may vary from 3 to 10% by volume when entering the contact system. These gases are available from ore roasting smelting and refining operations associated with nonferrous metals production.

Sulfur dioxide from metallurgical type operations is contaminated with dust and acid mist which must be removed prior to being introduced into the converter to prevent fouling of the catalyst. To remove impurities and excess water vapor, the gases are cooled to essentially atmospheric temperature.

Cleaning can be accomplished by the use of several different types of gas cleaning equipment, including cyclones, electrostatic precipitators, and scrubbers. Figure 19.2 is a block diagram showing an installation utilizing both wet and dry electrostatic precipitators for dust and mist removal.

The cleaned and cooled gases are passed through drying towers where they are scrubbed with concentrated sulfuric acid to remove the

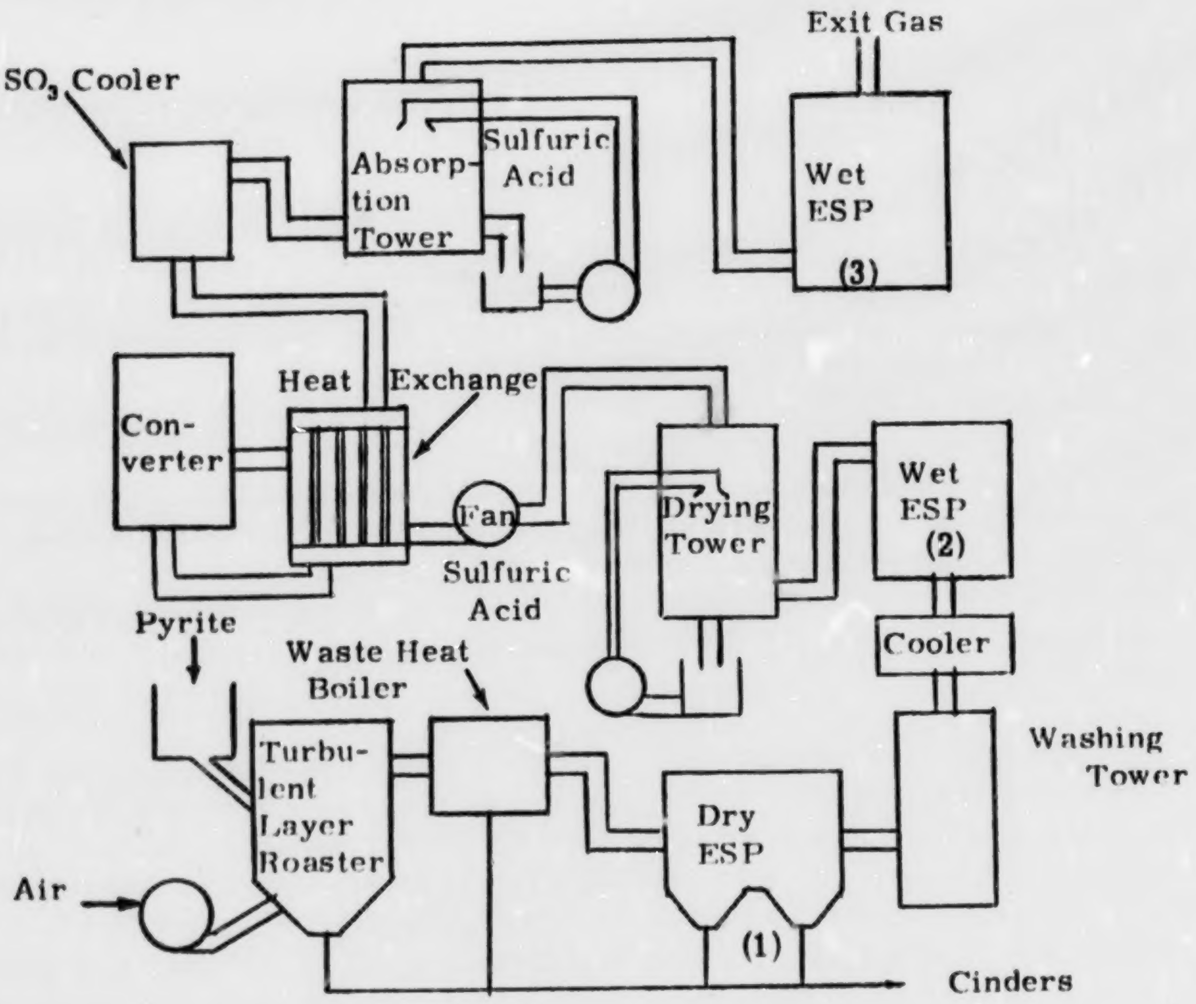

Exit Gas
SO_3 Cooler
Absorp-tion Tower
Sulfuric Acid
Wet ESP
(3)
Heat Exchange
Con-verter
Fan
Drying Tower
Wet ESP
(2)
Sulfuric Acid
Pyrite
Waste Heat Boiler
Cooler
Washing Tower
Turbu-lent Layer Roaster
Dry ESP
Air
(1)
Cinders

Figure 19.2. Applications for Wet and Dry Eletrostatic Precipitators in Contact Sulfuric Acid Plants Using Sulfur-Bearing Ores.

remaining water vapor. After cleaning, the gas is reheated to about 800°F by heat exchange with sulfur trioxide from the converter. The sulfur trioxide gas is thereby cooled to a suitable temperature for absorption in the final absorber tower.

Burning of hydrogen sulfide and spent-acid. Sulfuric acid is also produced by burning hydrogen sulfide and spent-acid. The raw materials may be either spent alkylation acid from petroleum refineries, hydrogen sulfide, or both. The effluent gas contains up to about 14% SO_2, and the temperature may be as high as 2400°F.

Gas cooling and heat recovery is accomplished in a waste heat boiler followed by heat exchangers. Mist removal is usually accomplished by electrostatic precipitation, and moisture removal by absorption in concentrated sulfuric acid. From the drying tower, the gas is introduced into a converter and treated in the same manner as gases from the ore roasting process.

In a few hydrogen sulfide or hydrogen sulfide plus elemental sulfur burning plants, the wet gases from the combustion process are charged directly into the converter after heat recovery, with no moisture or mist removal. The gas from the converter is then absorbed in concentrated sulfuric acid. In this "wet-gas" process, absorption is not highly efficient due to the excess moisture and acid mist content of the gas, and consequently requires a highly efficient mist recovery system following the absorber.

Converters and absorbers. In the catalytic converter, sulfur dioxide from any of the above sources reacts with excess oxygen from the combustion air in the presence of a solid catalyst material containing about 7% vanadium pentoxide to form sulfur trioxide. The oxidation reaction is exothermic and the temperature must be controlled by cooling between stages of the converter to maintain satisfactory equilibrium conditions. This cooling is achieved either by dried air injected into the converter, or by interstage heat exchangers.

Exit gas from the converter is cooled to approximately 450°F in a variety of heat recovery systems, after which it is sent to the absorber. In the absorber, the sulfur trioxide combines with water in 98 to 99%

sulfuric acid in the final recovery operation. In some plants, oleum (a solution of SO_3 in H_2SO_4) is produced. Additional absorption towers and heat exchangers are required for this operation.

Atmospheric emissions and control techniques. The primary atmospheric pollutant from the contact process is unreacted sulfur dioxide in the waste gas from the absorber since normal SO_2 to SO_3 conversion efficiencies range from 98 to 98.5%. In addition, unabsorbed SO_3 and H_2SO_4 mist from the reaction of SO_3 and small quantities of water vapor in the converter are present in the effluent gas. The acid mist, which forms when the temperature falls below the dew point, is of very small particle size and will pass through the absorber without being collected. Unless suitable collection equipment is provided, the mist is emitted from the process and constitutes an undesirable air pollutant.

The type of collection equipment needed to remove the acid mist depends to a great extent on the size of the particles in the absorber exit gas. Wire-mesh mist eliminators have collection efficiencies over 90% when most of the mist particles are of a diameter greater than 3 microns, which is the case when only 98% acid is being produced. When oleum is also produced, 85 to 95% of the mist particles leaving the oleum tower is less than 2 microns, and the wire-mesh pads do not effectively remove these fine particles.

Electrostatic precipitators have been used extensively for acid mist removal and are effective even for the small particle sizes. The pressure drop through the precipitator is less than 1 in. of water. High efficiency glass fiber mist eliminators are also used, even on plants where oleum is produced. To maintain high efficiencies, pressure drops on the order of 5-10 in. water are required. However, because of their low initial and maintenance costs, this type mist eliminator is finding increased application in sulfuric acid plants.

19.3 DESIGN AND OPERATING PARAMETERS FOR SULFURIC ACID MIST PRECIPITATORS

The standard design for an acid mist collector consists of a single-stage, vertical up-flow, pipe-type precipitator, as shown in Figure 19.3.

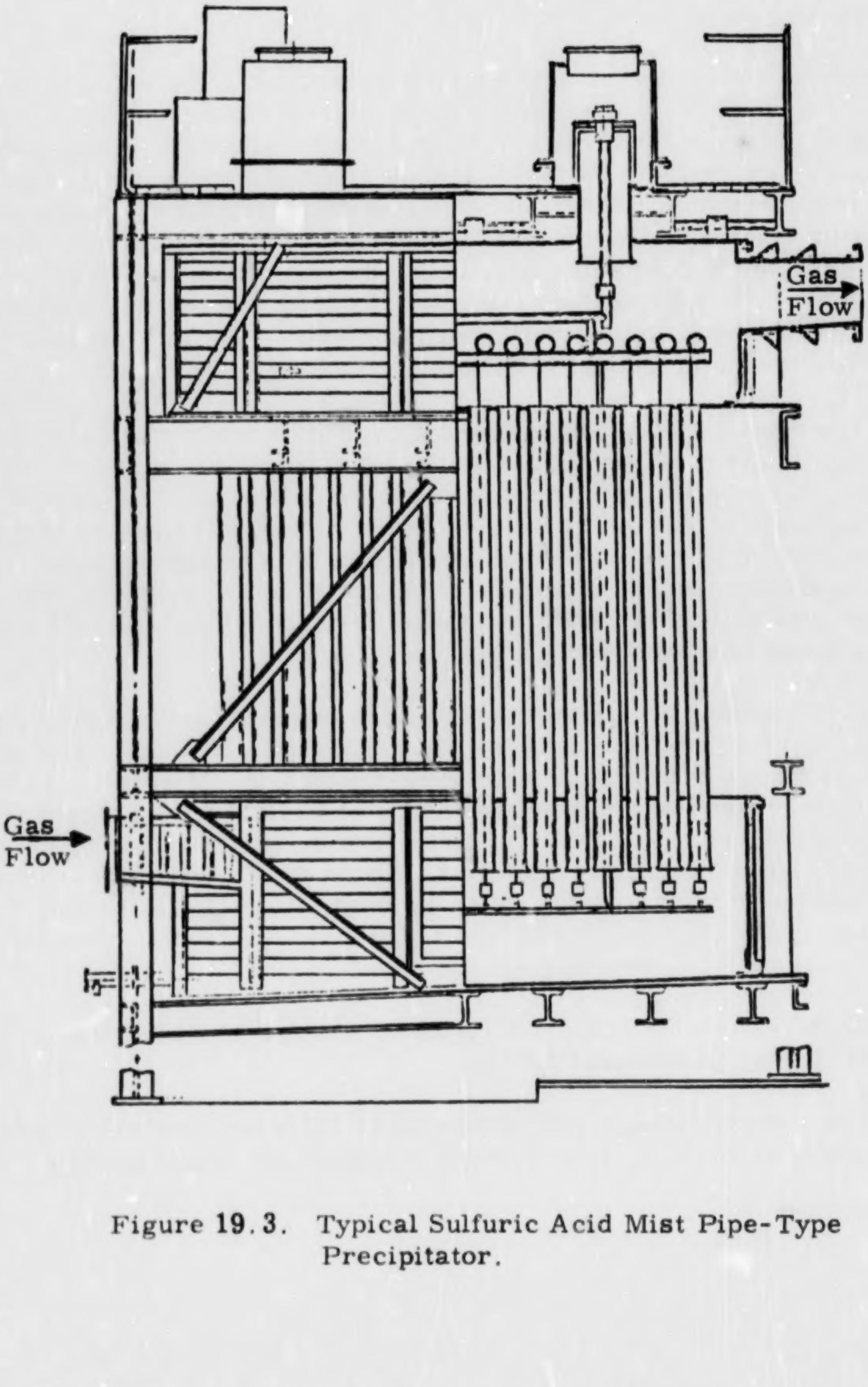

Gas
Flow
Gas
Flow

Figure 19.3. Typical Sulfuric Acid Mist Pipe-Type Precipitator.

The cylindrical shell is constructed of sheet lead supported by steel banding. Because of the material limitation, operating temperatures are limited to 180-200°F, with maximum pressures of about 20 inches water. Acid mist precipitators can also be made of steel, and precipitators of all steel construction have been in service for several years in plants producing acid of high concentration.

Discharge electrode system. Discharge electrodes can be constructed of lead with a cross section in the form of a four pointed star about $\frac{1}{2}$" across the points or they can be of steel construction.

Because of the possible shorting out of the high voltage system by acid, a specially designed discharge electrode support system is required. In a typical system, all members are constructed of, or shielded by, acid-resistant materials. The discharge electrode frame is supported by a beam which passes through the shell into insulator compartments located on each side near the top. There are two general compartment designs used. In the first, the support insulators are isolated from the interior of the precipitator by a liquid oil seal. Any acid condensing out on the precipitator side settles down through the oil to the bottom of the seal and flows to the outside through a liquid trap. In the second type, ambient air is forced through the seal to the inside of the precipitator in order to effect a blanket around the support insulator. The insulator is also protected against any acid that would drip off the compartment ceiling by a hood. In this design the acid accumulated on the bottom of the compartment also flows to the outside through a liquid trap.

Collecting electrode system. The collecting electrode system consists of pipes arranged in a circular grouping. They can be constructed of lead or steel. The electrodes proper are usually about 10 inches in diameter. The pipes are supported from a top header and are either completely contained in a precipitator shell or partially exposed, with an upper and lower distribution chamber, and collection sump.

Since the precipitate is free flowing, no electrode shaking or rapping is required. If cleaning is required, it is usually accomplished by washing with weak acid or water from sprays located in the top header.

Design parameters. Gas flow ratings for acid mist precipitators range from about 10,000 to 30,000 cfm at gas temperatures of 100°F to 180°F.

Collection efficiencies are usually specified at 95 to 99%, with the higher values predominant. To meet these requirements, the following basic design parameters are typically used.

Basic Parameter	Symbol	Range
Precipitation rate	w	0.20 - 0.30 ft/sec
Specific collection surface	$\frac{A}{v}$	200 - 400 sq ft/1000 cfm
Pipe diameter	d	10 in.
Gas velocity	v	3 - 6 ft/sec
Corona power	$\frac{P_c}{V}$	100 - 500 watts/1000 cfm
Corona power density	$\frac{P_c}{A}$	0.5 - 2.0 watts/sq ft
Rectifier voltage	kVp	75 - 100 kV peak

Inasmuch as sulfuric acid is highly conductive electrically, there is no resistivity problem to disturb the design. Corona current suppression can occur owing to the high concentration of submicron particles, and this can influence the corona power and rectifier design.

Operating parameters. Data on process and operating parameters are summarized in histogram form in Figures 19.4 through 19.8 for the years 1945-1969. The histograms show the percent distribution of each parameter in terms of total gas flow for the 25 year period and include gas temperature, acid mist concentration, precipitator gas velocity, precipitator corona power, and precipitator field strength.

General observations and discussion of trends--sulfuric acid mist precipitators. Data indicate that sulfuric acid mist precipitators are being constructed in larger unit sizes to treat acid mist gases from larger plants. Average precipitator size has nearly tripled in the past 20 years, from about 10,000 cfm in 1944-1949 to about 30,000 cfm in 1964-1969. Further, the annual installed capacity is growing at a significant rate with

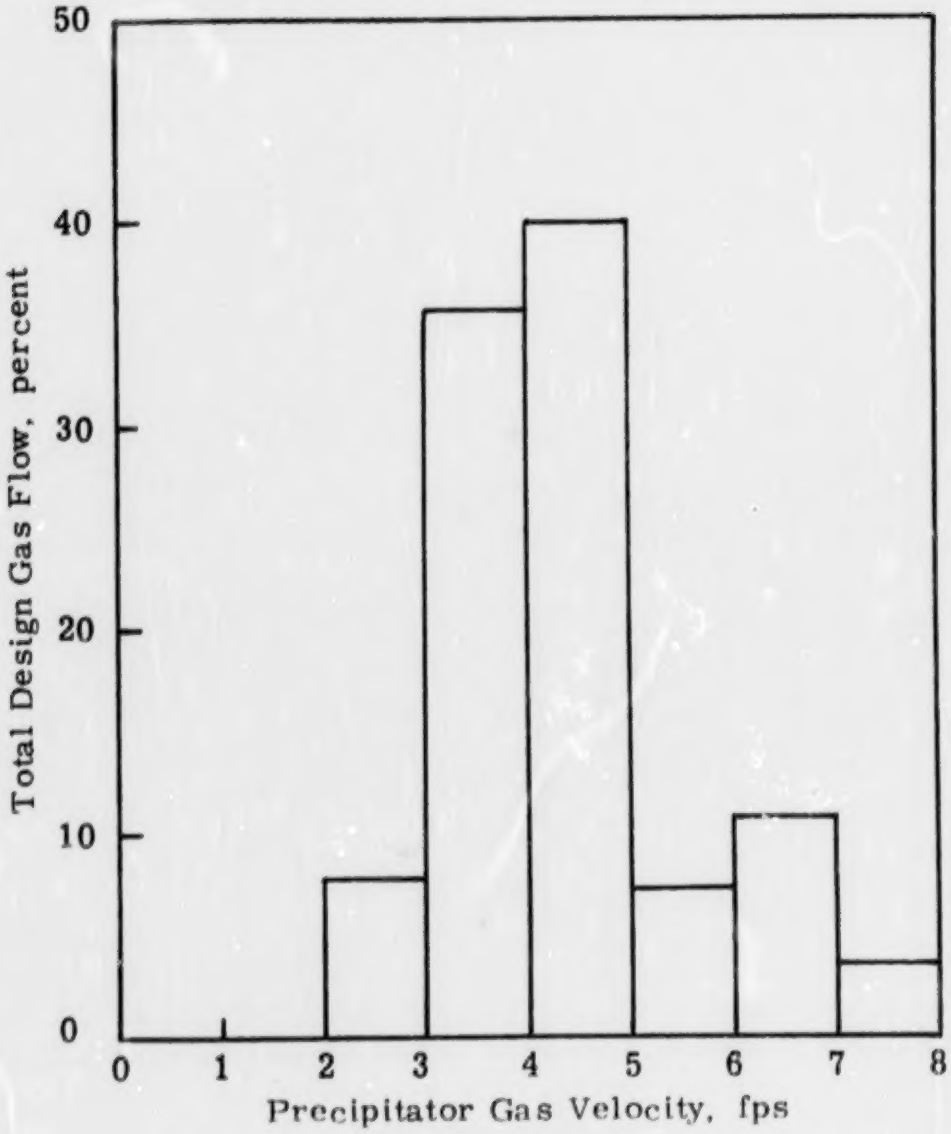


Figure 19.4. Distribution of Sulfuric Acid Mist Precipitator Gas Velocity, 1945-1969.

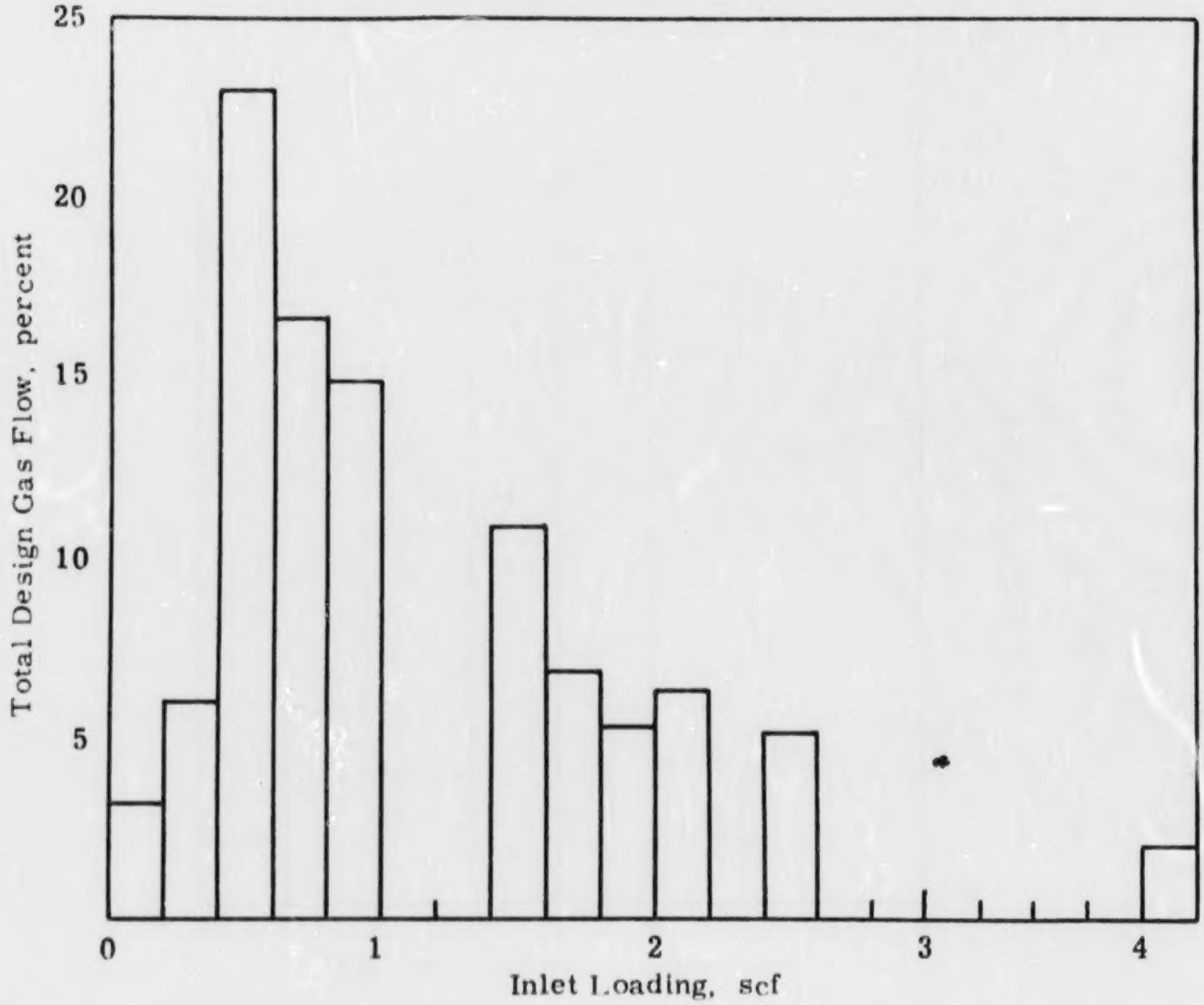


Figure 19.5. Distribution of Sulfuric Acid Mist Precipitator Inlet Mist Loading, 1945-1969.

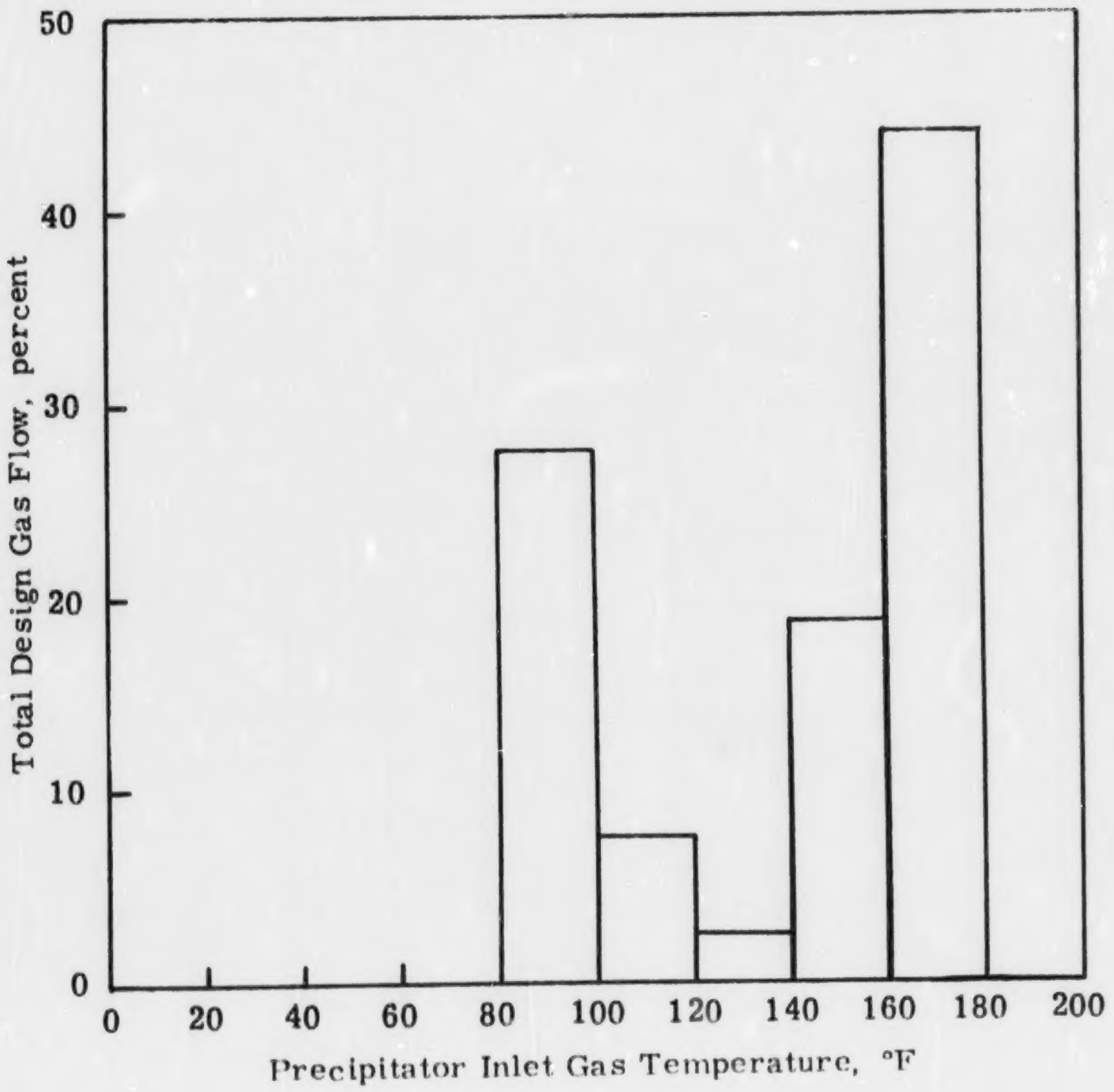


Figure 19.6. Distribution of Sulfuric Acid Mist Precipitator Inlet Temperature, 1945-1969.

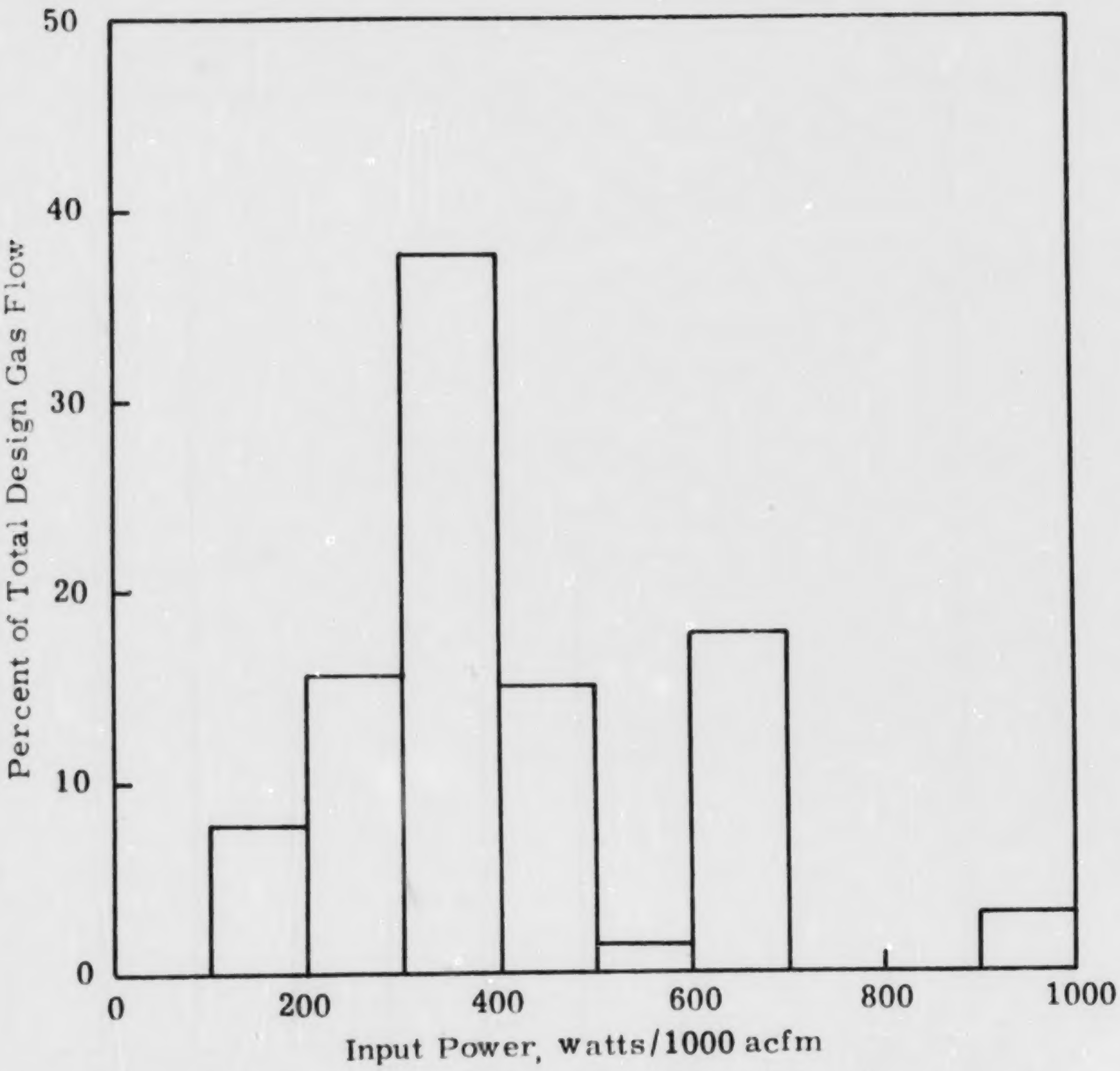


Figure 19.7. Distribution of Sulfuric Acid Mist Precipitator Input Power, 1945-1969.

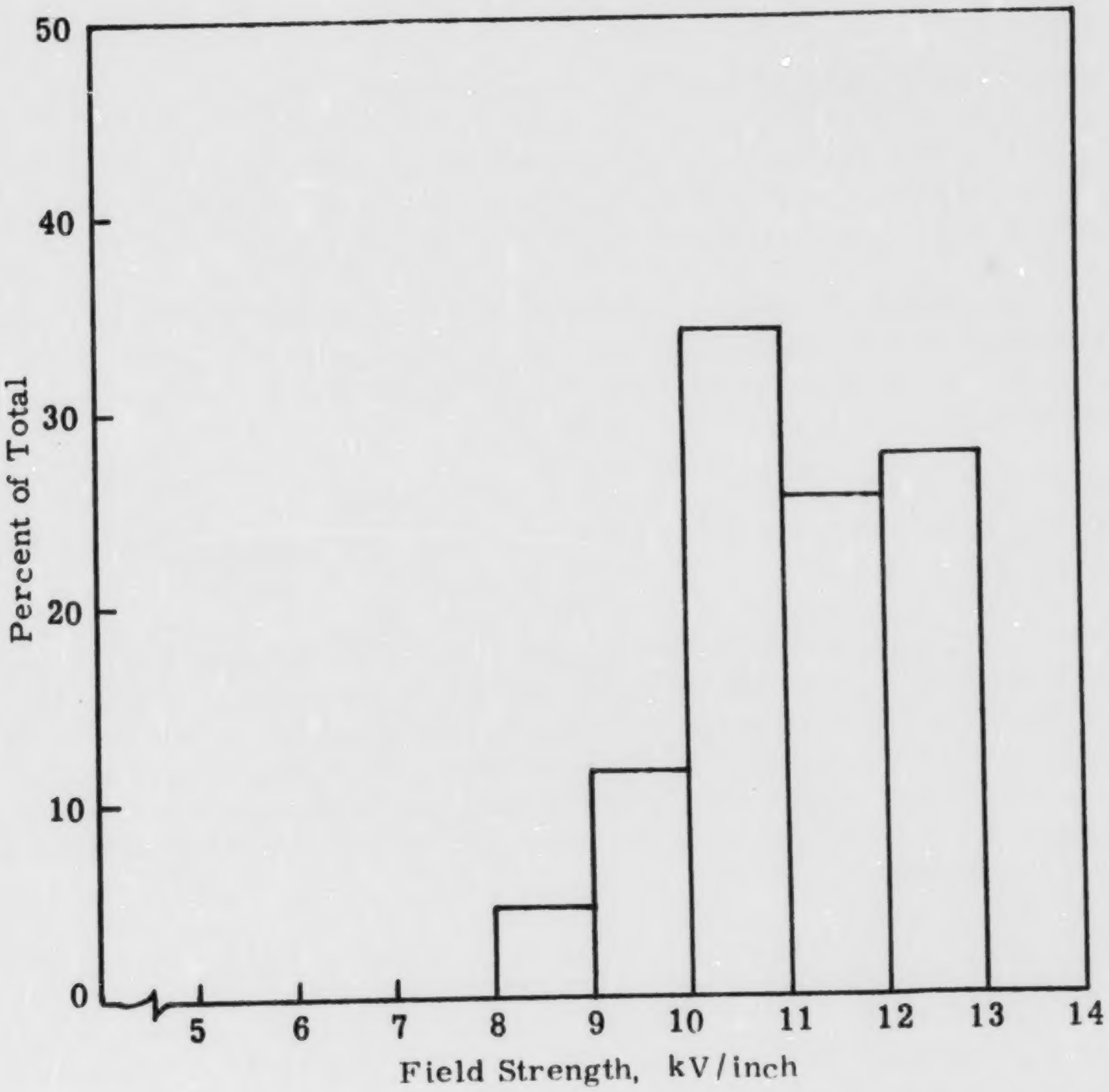


Figure 19.8. Distribution of Sulfuric Acid Mist Precipitator Field Strength, 1945-1969.

the total capacity nearly doubling in the past six years.

There is a general trend to higher design efficiency. Despite this trend, the installed cost/acfm of these precipitators has decreased over the last decade, partly because of an increase in precipitator average size.

A major factor in the cost of these precipitators is the material of construction and the fabrication technique associated with lead. There is activity in the industry to develop suitable nonmetallic precipitators for this application.

19.4 SULFURIC ACID MIST PRECIPITATOR INSTALLATIONS AND COST DATA

The total number of sulfuric acid mist precipitators installed 1945-1969 is estimated to be 120, with a total gas flow rating of 2,230,000 cfm. Table 19.1 summarizes these installations, while cumulative and yearly average cfm data are shown in Figures 19.9 and 19.10, respectively. Figure 19.11 depicts the trend toward larger precipitator sizes and Figure 19.12 the increase in design efficiencies which has occurred particularly in the past decade.

Cost data. Precipitator cost data on an installed basis are presented in terms of cost versus gas flow rate for several efficiency levels for the period 1960 to 1969. Limited data on FOB precipitator cost are included. The data are summarized in Table 19.2 and Figure 19.13. The spread in the data is indicated on the figure. The scatter in cost data, at a given efficiency and gas rate, can be attributed to many factors, the most important ones being the size of the precipitator as related to mist characteristics, primarily particle size, and gas conditions; the geographical location of the installation (particularly for erected cost since labor costs can vary considerably across the country); the pricing-profit policy of various corporations bidding on the job; whether the installation is a backfit or upgrading of an existing installation which may require additional improving to "shoehorn" the precipitator into the over-all installation; and the type and degree of sophistication of the electrical system. Table 19.3 shows the trend in average installed cost of sulfuric acid mist precipitators over the period 1945-1969.

Table 19.1

Sulfuric Acid Mist Precipitator Installations 1945-1969

Pptr. Contract Year	Pptr. Capacity acfm	No. of Install.	No. of Pptrs.	Cumulative Capacity acfm	Avg. Capacity Annual Capacity	acfm Precipitator Basis
1945	29, 200	2	3	29, 200	5, 840	9, 733
1946	-	-	-	-		
1947	-	-	-	-		
1948	-	-	-	-		
1949	-	-	-	-		
1950	71, 050	3	6	100, 250	95, 000	13, 970
1951	89, 200	3	6	189, 450		
1952	62, 800	3	5	252, 250		
1953	39, 000	2	3	291, 250		
1954	212, 950	6	14	504, 200		
1955	158, 750	8	12	662, 950	94, 106	13, 070
1956	101, 700	5	7	764, 650		
1957	147, 540	7	12	912, 190		
1958	46, 800	1	2	958, 990		
1959	15, 740	3	3	974, 730		
1960	21, 080	2	2	995, 810	35, 976	16, 353
1961	9, 400	1	1	1005, 210		
1962	-	-	-	-		
1963	149, 400	3	8	1154, 610		
1964	-	-	-	-		
1965	79, 050	3	4	1233, 660	215, 594	29, 943
1966	167, 900	3	9	1401, 560		
1967	183, 120	3	9	1584, 680		
1968	198, 700	3	6	1783, 380		
1969	449, 200	4	8	2232, 580		
Total			120			

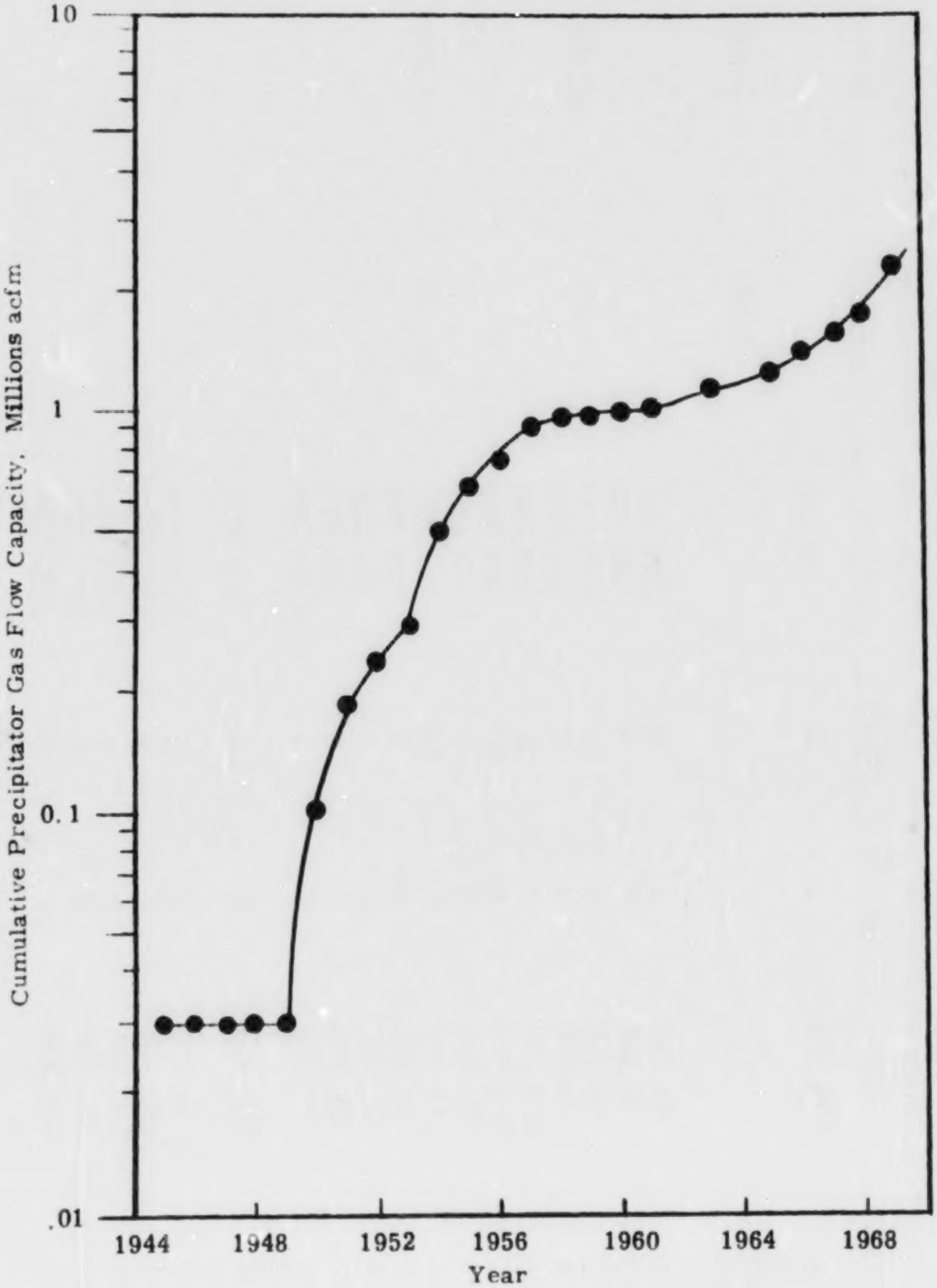


Figure 19.9. Cumulative Gas Flow Capacity of Sulfuric Acid Mist Precipitators, 1945-1969.

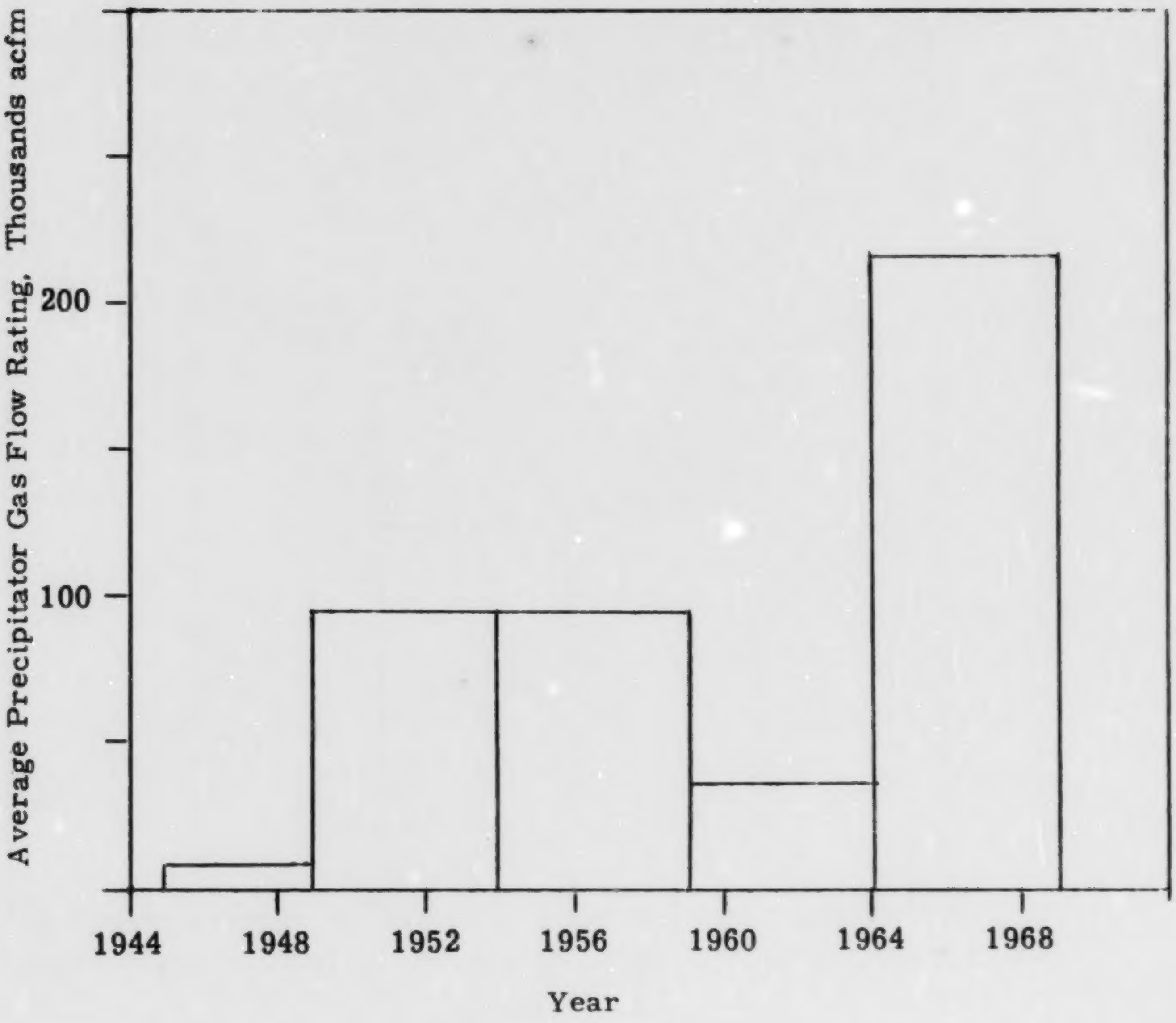


Figure 19.10. Yearly Average Gas Flow Ratings of Sulfuric Acid Precipitators; Data in Five-Year Intervals, 1945-1969.

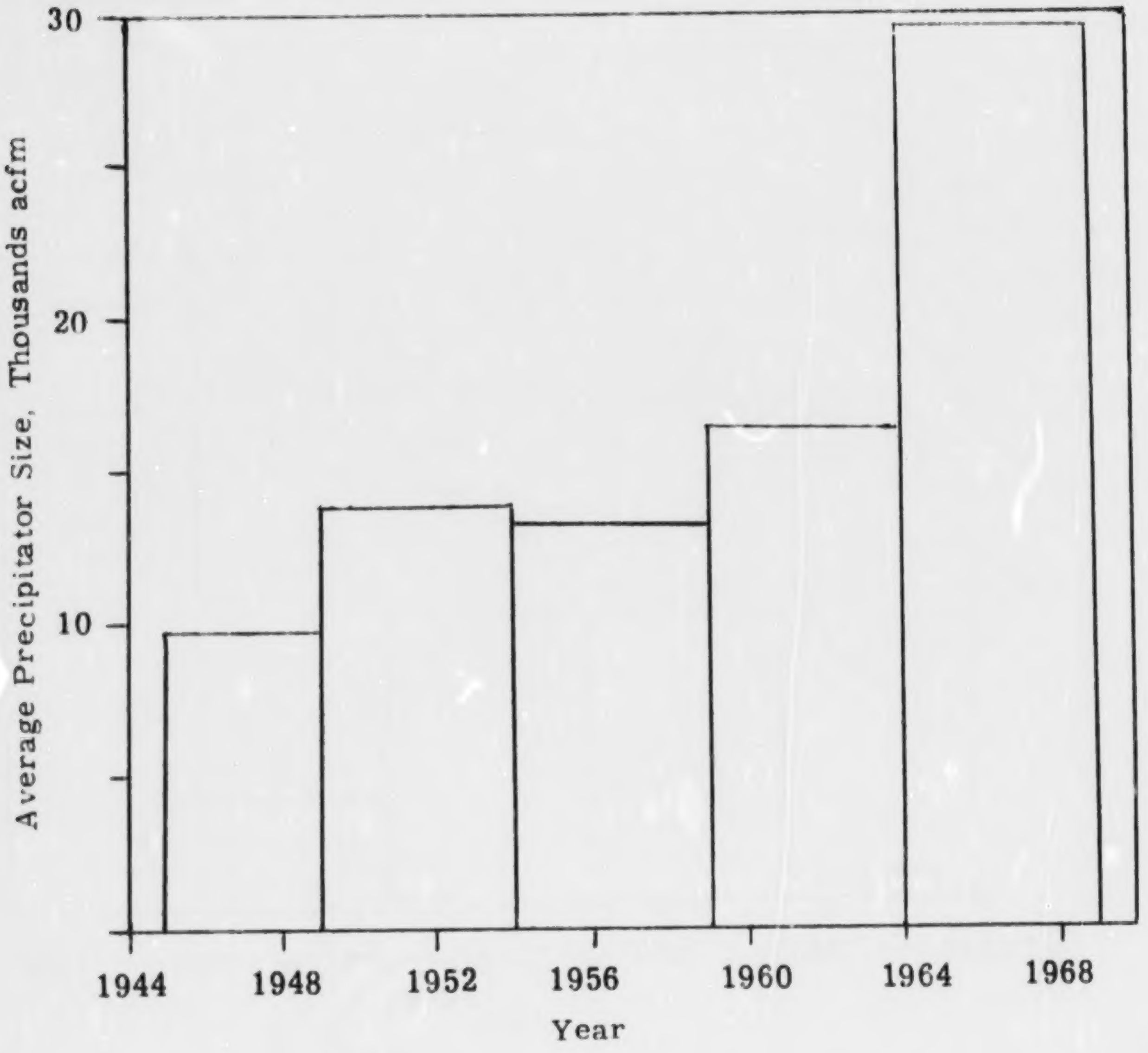


Figure 19.11. Average Size of Acid Mist Precipitators in Terms of Gas Flow Capacity; Data in Five-Year Intervals, 1945-1969.

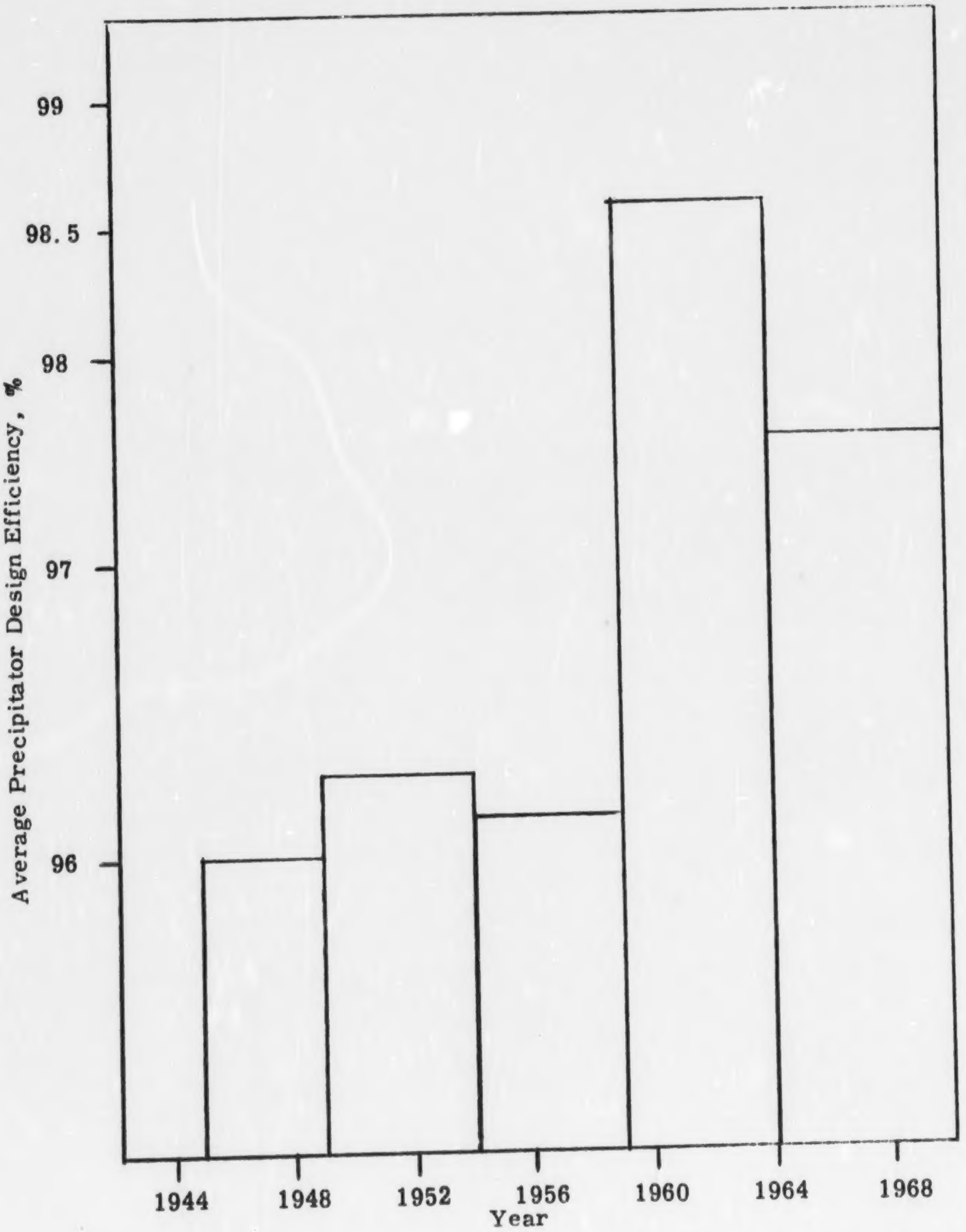


Figure 19.12. Design Efficiency Trend for Sulfuric Acid Mist Precipitators Weighted on acfm Basis, 1945-1969.

Table 19. 2

Summary of Sulfuric Acid Mist Precipitator Costs 1960-1969

	0 - 10, 000 cfm				10, 000 - 25, 000 cfm				25, 000 - 50, 000 cfm			
	1965-1969		1960-1965		1965-1969		1960-1965		1965-1969		1960-1965	
Efficiency Range	FOB*	Erected*	FOB	Erected	FOB	Erected	FOB	Erected	FOB	Erected	FOB	Erected
90 - 95	-	-	-	-	-	-	-	-	-	2. 21 (1)	-	-
95 - 99	-	-	14. 5 (1)	6. 00 (1)	3. 77 (1)	5. 84 (4)	-	-	-	4. 65 (1)	-	3. 94 (2)
99+	-	-	-	-	-	-	-	-	-	-	-	-

	50, 000 - 100, 000 cfm				> 100, 000 cfm			
	1965-1969		1960-1965		1965-1969		1960-1965	
Efficiency Range	FOB	Erected	FOB	Erected	FOB	Erected	FOB	Erected
90 - 95	-	-	-	-	-	-	-	-
95 - 99	-	6. 22 (1)	2. 93 (1)	-	-	2. 53 (5)	-	-
99+	-	3. 71 (3)	-	-	-	-	-	-

* Note: Costs are \$/acfm. Number in parentheses is number of installations on which contract prices were averaged.

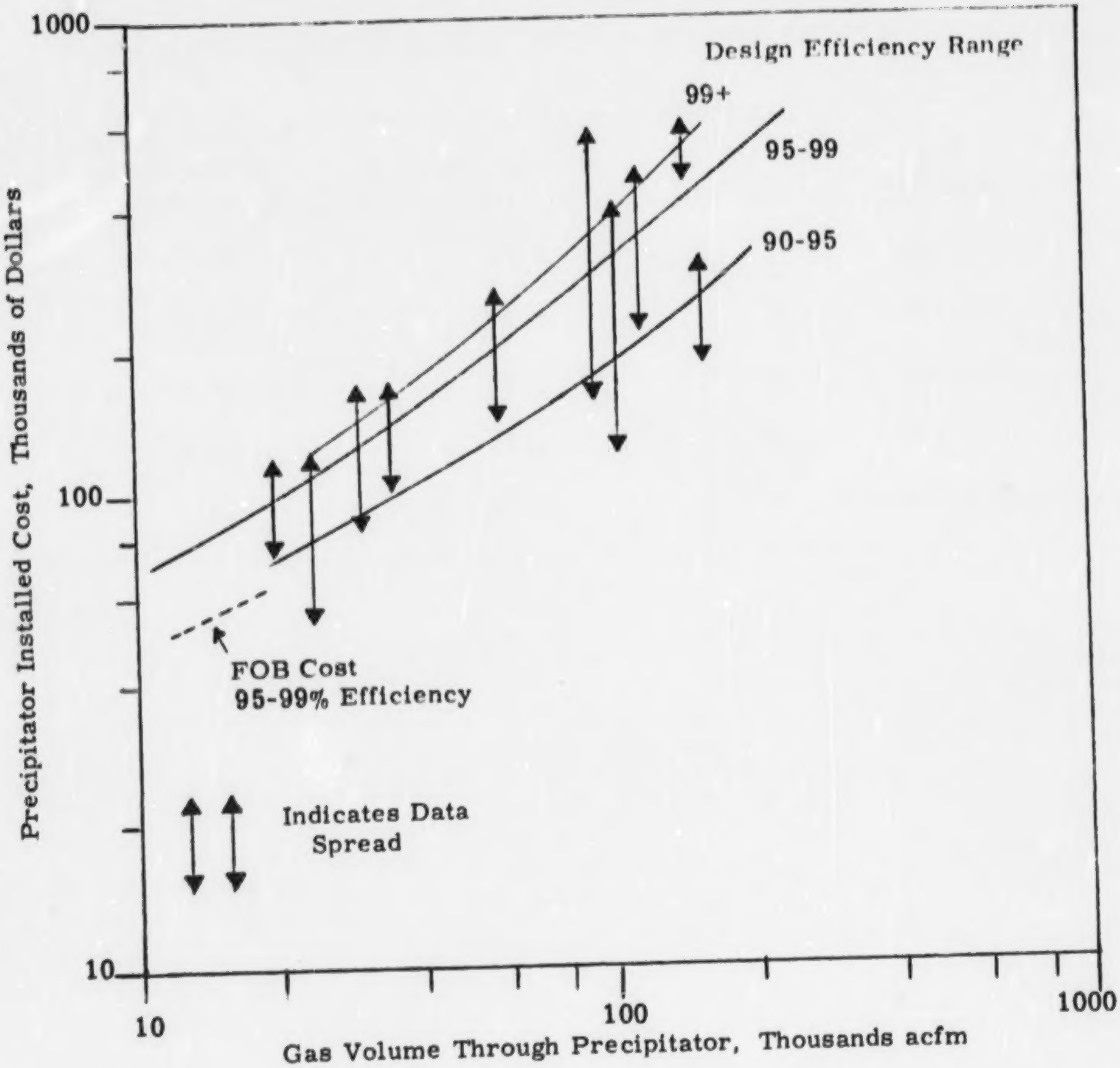


Figure 19.13. Installed Cost of Sulfuric Acid Mist Precipitators, 1965-1969.

Table 19.3

Average Cost per acfm for Sulfuric Acid Mist Precipitators

Contract Period	No. of Installations	Average Installed Cost $/acfm	Weighted Design Efficiency % (acfm Basis)
1945-1949	2	3.82	96
1950-1954	11	5.18	96.4
1955-1959	16	5.62	96.2
1960-1964	3	4.16	98.6
1965-1969	15	3.76	97.7

Operating costs. Operating costs for a 65,000 cfm sulfuric acid mist precipitator are as follows.

Precipitator FOB Cost	$115,000	(1.70/cfm)
Precipitator Erected Cost	280,000	(4.20/cfm)
Operating Cost*	2,900	
Maintenance Costs	1,865	
Capital Costs ($12\frac{1}{2}$% of Instl Costs)	35,000	
Total Yearly Operating Costs	$39,765	

Electrostatic precipitators on sulfuric acid plants using roaster exhaust gas. In contact sulfuric plants, in which sulfur-bearing ores are the sources of sulfur dioxide, there are three separate applications of precipitators for gas cleaning, as indicated in Figure 19.2. These are: (1) removal of the dry dust and acid fume from the hot gases (400-800°F) after they leave the furnace and before the acid scrubbers; (2) removal of the sulfuric acid mist and solid impurities from the cooled gases (90-100°F) after the gases have passed through the acid scrubbers and prior to entering the converter; and (3) cleaning of tail gases after the absorber. The latter applications involve the purification of gases for process reasons as opposed to control of air pollution.

The precipitators used to clean the hot gases and remove dry dust and fume are usually of the plate type and constructed of mild steel.

Limited data are available on the design and performance of precipitators in process gas cleaning. Table 19.4 lists some of the design and performance data on a single precipitator installation for mist and solids removal ahead of the converter.

19.5 PRECIPITATORS FOR THE ELEMENTAL PHOSPHOROUS INDUSTRY

The areas of interest in the phosphorous industry, as far as electrostatic precipitators are concerned, are the reduction of phosphate rock to produce elemental phosphorus and the production of phosphoric acid.

Elemental phosphorus industry. Elemental phosphorus is produced by the reduction of prepared phosphate rock in a furnace (usually an electric-

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* Labor, Supervision, and Overhead based on Records.

Table 19.4

Comparison Between Design and Performance Parameters for An ESP Operating on Acid Mist from a Sulfuric Acid Plant

Parameter	Design	Performance
Gas flow (acfm)	67,300	62,300
Efficiency (%)	99	99.3
Area (sq ft)	23,538	23,538
Collection voltage (kV)		66
Corona current (mA)		300
Power (kW)		20
FOB cost ($)	115,230	
Installed cost ($)	284,000	
A/V ratio (ft^2/cfm)	0.35	0.378
V/A ratio (cfm/ft^2)	0.0475	0.044
Precipitation rate parameter w (ft/sec)	0.219	0.202
(cm/sec)	6.7	6.2
FOB cost ($/acfm)	1.72	
Installed cost ($/acfm)	4.20	
Installed cost ($/sq ft)	12.00	

arc furnace). In 1965, about 1.2 billion pounds of elemental phosphorus was produced in the United States. Several manufacturers employ the electrothermic reduction process for manufacture of phosphorus. A major source of information on this subject is the Tennessee Valley Authority[1] which made significant contributions to the technology of the process.

The furnace is charged with phosphate rock (fluorapatite), silica, and coke. Temperatures of 2300-2700°F are maintained in the furnace, and at these temperatures, silica reacts with the phosphate rock to liberate P_2O_5, which is reduced to elemental phosphorus by the carbon. The reaction is as follows:

$$CaF_2 \cdot 3Ca_3(PO_4)_2 + 9SiO_2 + 15C \rightarrow CaF_2 + 9CaOSiO_2 + 6P\uparrow + 15CO\uparrow$$

The carbon for the above reaction is usually coke, but anthracite has also been used. The silica is needed to form a low melting slag that serves as a fluid medium for the reduction reaction. The reaction in the furnace is exothermic, but additional heat is required to maintain the reaction temperature in the furnace. Three carbon electrodes, connected to a 3-phase power supply, provide this energy. The gaseous effluent from the furnace (70.4% CO by weight, 0.2% CO_2, 27.2% P_4, 0.4% N_2, 0.2% H_2, and 1.6% dust)[2] is cleaned with an electrostatic precipitator operating above the condensation temperature of the phosphorus vapor (525-620°F).[3] The dust (5 to 20 gr/scf) consists of fine particles from the burden and from the furnace reactions. If not removed, this dust would be collected in the phosphorus during its condensation and would lower the purity of the liquid phosphorus product. Dust can also be removed from the condensed product by centrifugation, decantation, and filtration of the molten phosphorus. However, since filtration is quite difficult, it is used mainly for final purification of the product, and electrostatic precipitators are used prior to condensation to remove the major portion of the dust. In this application, the precipitator is a part of the plant production facility as opposed to an air pollution control device.

Precipitators typically remove 90 to 99% of the dust from the gases. The major problem encountered is the build-up of dust on the inside walls of the collector tubes. This dust adheres strongly and is not completely removed by rapping. The unit must be shut down periodically to ream out the tubes. Very little dust adheres to the emitting wires, and occasional wire breakage is usually due to fatigue from rapping.

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[1] Refer to the bibliography for this chapter.

The inside surfaces of the precipitator must be heated to prevent condensation of phosphorus. This is accomplished either by electric heaters or by passing hot inert gas through a slightly pressurized jacket. The latter method ensures that leaks in the precipitator walls do not allow air to enter the precipitator. Oxygen in the presence of carbon monoxide and phosphorus vapor would cause damaging combustion in the precipitator.

The dust collected in the precipitator is accumulated in a hopper where it is removed from the system to an external dust bin. Removal is accomplished through an air-lock valve arrangement. This dust contains considerable phosphate and is sometimes used as fertilizer. Some manufacturers remove the dust from the bins with a water spray, after which the slurry is dried, briquetted, and returned to the furnace to recover the phosphorus.

The gases leaving the precipitator pass through a water spray tower maintained at 125° F where the phosphorus vapor condenses. The water dissolves gaseous fluosilicates and becomes acidic. Soda ash or ammonia is often added to reduce the acidity when this scrubber water is recirculated.

Carbon monoxide is the main component of the gases leaving the spray tower. Some plants burn this gas to provide heat for other parts of the process; others flare the carbon monoxide to dispose of it.

Two molten by-products, ferrophosphorus and slag, are removed from the bottom of the furnace. Ferrophosphorus is sold for use in steelmaking, and slag is used as fill material for construction.

Figure 19.14 shows a typical elemental phosphorus plant flow diagram.

Description of electrostatic precipitator and subsystems. The precipitator is designed for vertical gas-flow as a single-stage pipe type unit (Figure 19.15). The pipes are usually about 11 inches in diameter and 16 feet long. In order to achieve the high efficiency required, two sections are used in series. Since the precipitator must be preheated, the collecting pipes are jacketed for hot gas circulation. Electric heaters are provided around the pipe in order to maintain the necessary operating temperature.

Discharge electrodes are generally vertical ($\frac{3}{16}$ or $\frac{1}{4}$") square bars spanning the full height of the collecting electrode. They are suspended

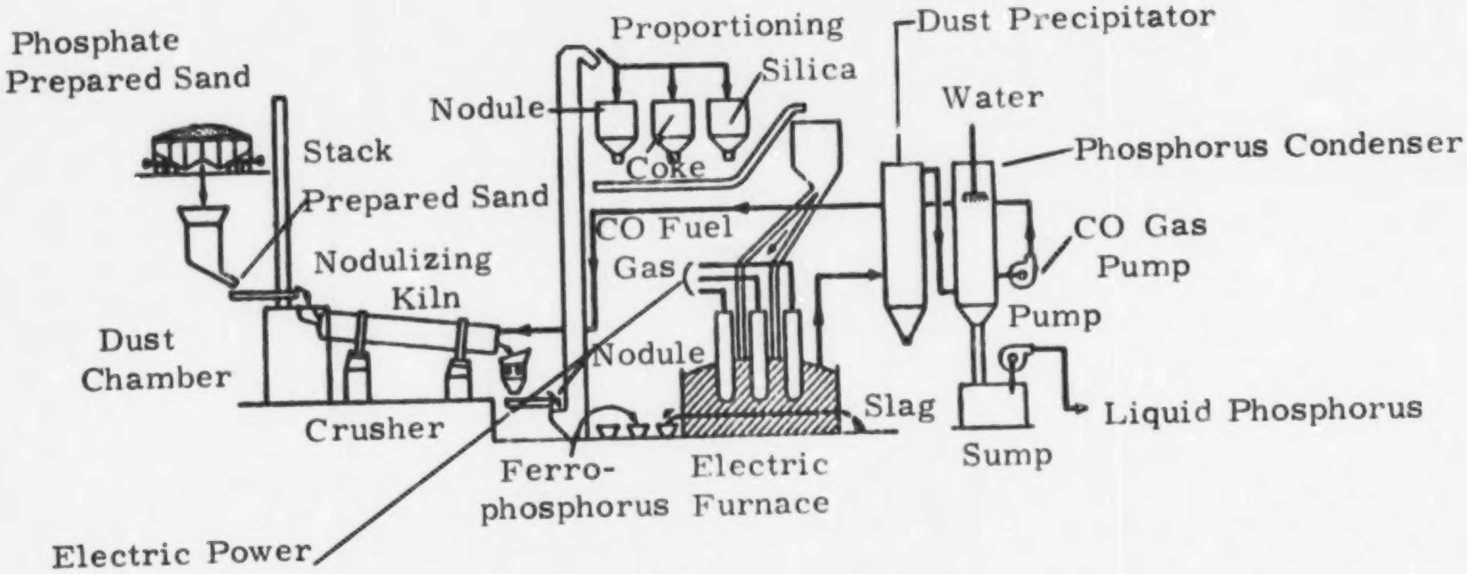
Phosphate
Prepared Sand
Stack
Prepared Sand
Nodulizing
Kiln
Dust
Chamber
Crusher
Electric Power
Proportioning
Nodule
Silica
Coke
CO Fuel
Gas
Nodule
Ferro-
phosphorus
Electric
Furnace
Slag
Dust Precipitator
Water
Phosphorus Condenser
CO Gas
Pump
Pump
Liquid Phosphorus
Sump

Figure 19.14. Elemental Phosphorus Production by the Electric Arc Furnace Process.

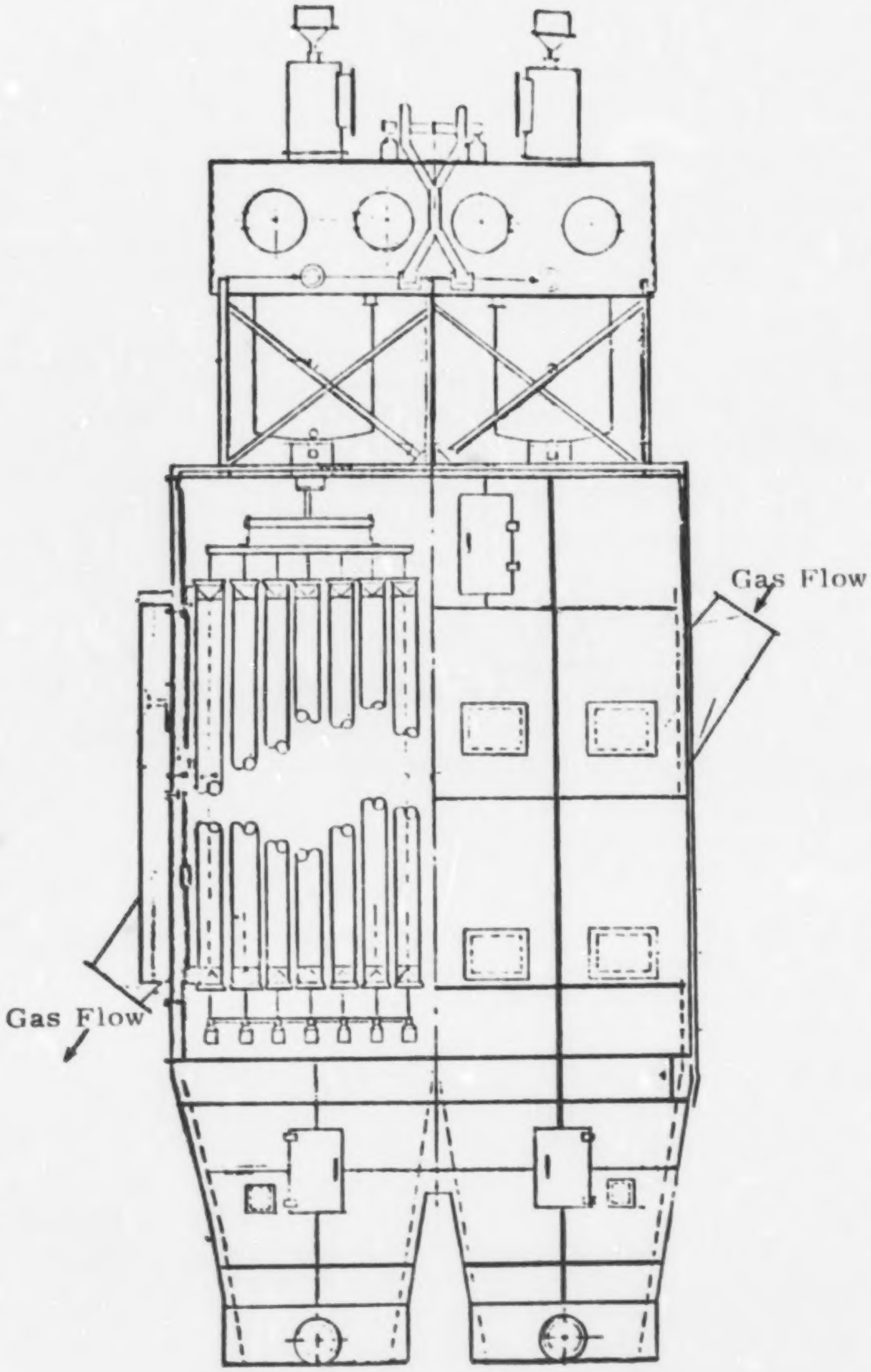
Gas Flow
Gas Flow

Figure 19.15. Typical Electric Furnace Phosphorus Precipitator.

from a support frame at the top, and held taut by weights at the bottom. Steadying frames and guides hold the wires in place.

The cylindrical collecting electrodes are usually constructed of steel pipe flared into square sections at the upper end and welded to form a gas tight bundle.

Since the precipitator operating temperature is relatively high, the hot collecting electrode pipes require gentle cleaning action to avoid damaging them. The reciprocating bumper mechanism generally used for this cleaning action consists of rapper bars running through the precipitator between the pipes and protruding out of the shell. Cross arms attached to the bars are provided with hammer heads. By moving the bars back and forth the hammers are made to strike anvils attached to the pipes. The rapper bars are attached to manual levers or an automatic mechanism on the outside of the precipitator shell. Operation is usually periodic and can be arranged for manual or automatic operation.

A form of deflector baffle is used in the hoppers of vertical up-flow precipitators for control of gas distribution. Where the gas enters the precipitator in a horizontal direction at hopper level, baffles are installed at strategic points in order to deflect the gas upward in such a fashion as to maintain good flow distribution. The baffles are usually hung from the bottom of the collecting electrodes or the lower header sheet in a pipe-type design. In some pan-hopper designs, the baffles are supported from the floor as hurdles.

A special kind of hopper and dust removal system must be used in connection with a precipitator employed for cleaning the gases emitted from a phosphorus-producing furnace. It is usually circular in cross section and contains a continuously operating dust plow or mixer to prevent congealing of the collected material. It is usually provided with heaters to prevent phosphorus condensation. Since the collected dust contains some condensed and absorbed elemental phosphorus which ignites when exposed to air, the dust must be kept out of contact with air.

An air tight dust bin is located directly below the hopper. The bin is connected to the hopper by an air-lock valve mechanism through which dust is periodically transferred from hopper to bin. A screw conveyor, which transports the dust through a sealed external section to closed transport containers, is located in the bottom of the bin.

It is estimated that about 20 precipitators for hot phosphorus applications have been contracted for during the period 1938-1969, with a total rated capacity of 165,300 acfm. Table 19.5 is a summary of hot phosphorus precipitator installations showing annual number of installations, annual number of precipitators, and annual design capacity for the period 1938-1969. The four-year average precipitator capacity is also shown, as is the average precipitator size. Figure 19.16 shows the cumulative capacity of elemental phosphorus precipitators during the period 1938-1969. Figure 19.17 shows the yearly and four-year average installed capacities in terms of gas volumes handled. Figure 19.18 shows the average precipitator capacity during four-year intervals from 1938-1969. A summary of design efficiencies for these precipitators is given in Table 19.6. Efficiencies are weighted on an acfm basis; four-year average efficiencies are also reported.

Precipitator applications. The precipitator gas velocity, inlet gas temperature and inlet particulate loading are summarized for the years from 1938 through 1969 in Table 19.7. The number of installations and total gas volumes involved in the statistical analysis are presented. No data are available on power input or field strength. Figures 19.19 to 19.21 show the range of gas velocities, inlet gas temperatures, and dust loadings.

Because of the difficulty in testing precipitators in this application, no test data are available. Table 19.8 summarizes the design data on two installations and shows the electrical energization equipment and precipitator size.

Cost data. Precipitator costs, both FOB and erected, for electric furnace phosphorus installations are presented in Table 19.9 for 10-year periods between 1938 and 1969. Due to the limited data, no graphical presentation is included. Data are presented as [illegible]ction of gas volume range and design efficiency.

Table 19.10 is a summary of four-year average precipitator size, weighted precipitator efficiency and average precipitator cost, dollars/acfm both on FOB and erected basis.

General observations and discussion of trends - electric furnace phosphorus precipitator. Continued growth of electric furnace phosphorus capacity is foreseen, with added capacity installed in large increments. Treatment of the furnace off-gas requires exposure to temperature in

Table 19.5

Hot Phosphorus Precipitator Installations 1938-1969

Pptr. Contract Year	No. of Install.	No. of Pptrs.	Pptr. Capacity acfm	Cumulative Capacity acfm	Average Capacity acfm	
					4-year (a) Average	Pptr. (b) Basis
1938	1	1	260	260		
1939	2	2	2,650	2,950		
1940	0	0	0	2,950	1,728	1,728
1941	1	1	4,000	6,910		
1942	1	1	5,000	11,910		
1943	0	0	0	11,910		
1944	2	2	6,500	18,410	3,458	3,458
1945	1	1	2,330	20,740		
1946	0	0	0	20,740		
1947	0	0	0	20,740		
1948	0	0	0	20,740	875	3,500
1949	1	1	3,500	24,240		
1950	0	0	0	24,240		
1951	0	0	0	24,240		
1952	2	2	13,300	37,540	5,575	7,433
1953	1	1	9,000	46,540		
1954	0	0	0	46,540		
1955	1	1	4,300	50,840		
1956	0	0	0	50,840	1,075	4,300
1957	0	0	0	50,840		
1958	1	1	12,200	63,040		
1959	0	0	0	63,040		
1960	0	0	0	63,040	8,100	10,800
1961	2	2	20,200	83,240		
1962	0	0	0	83,240		
1963	1	1	11,400	94,640		
1964	1	1	12,000	106,640	5,850	11,700
1965	0	0	0	106,640		
1966	2	2	32,050	138,690		
1967	2	2	26,600	165,290		
1968	0	0	0	165,290	14,662	14,662
1969	0	0	0	165,290		

(a) Pptr. capacity (acfm) divided by number of years in period.
(b) Pptr. capacity (acfm) divided by number of precipitators installed in period.

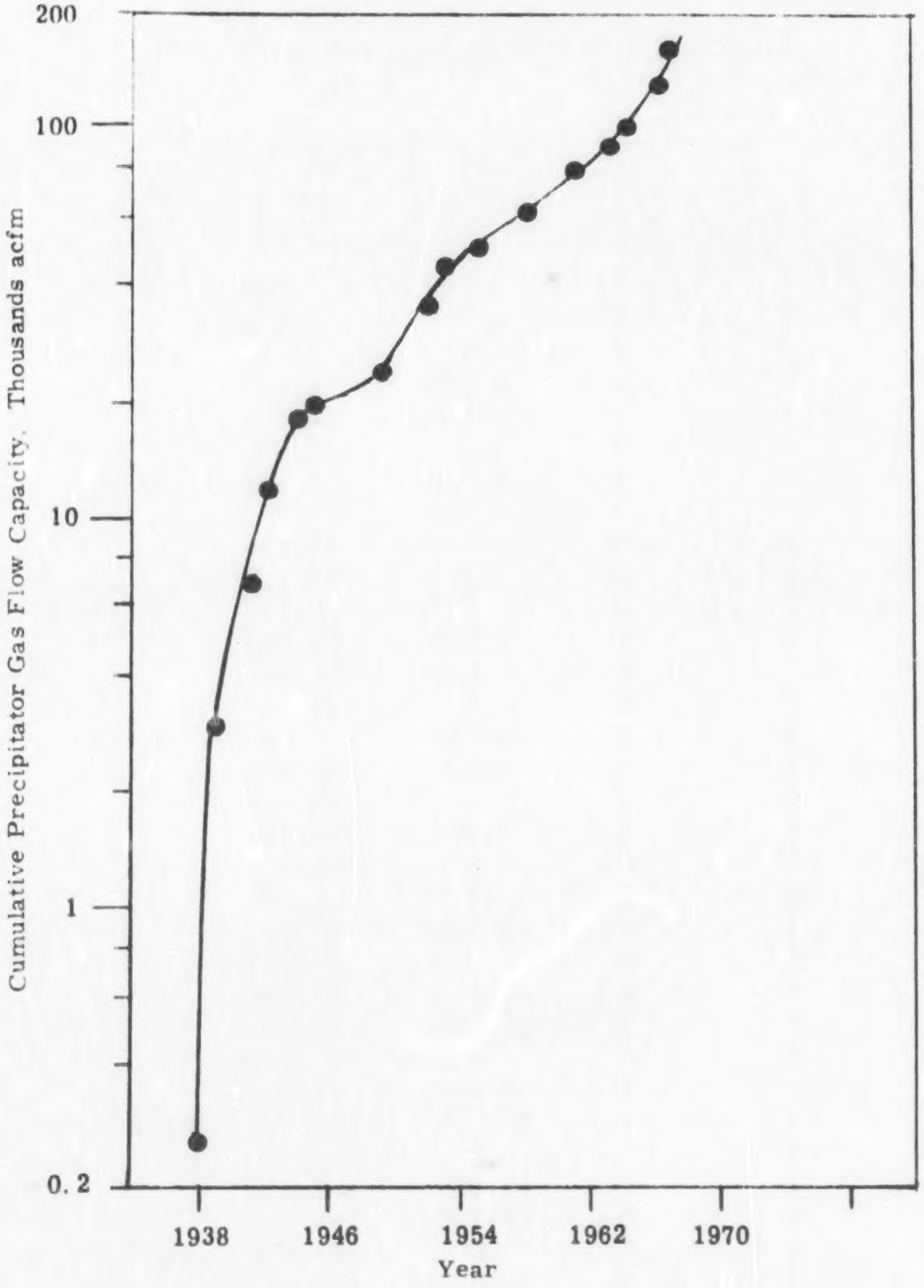

Figure 19.16. Cumulative Gas Flow Capacity of Elemental Phosphorus Precipitators Since 1938.

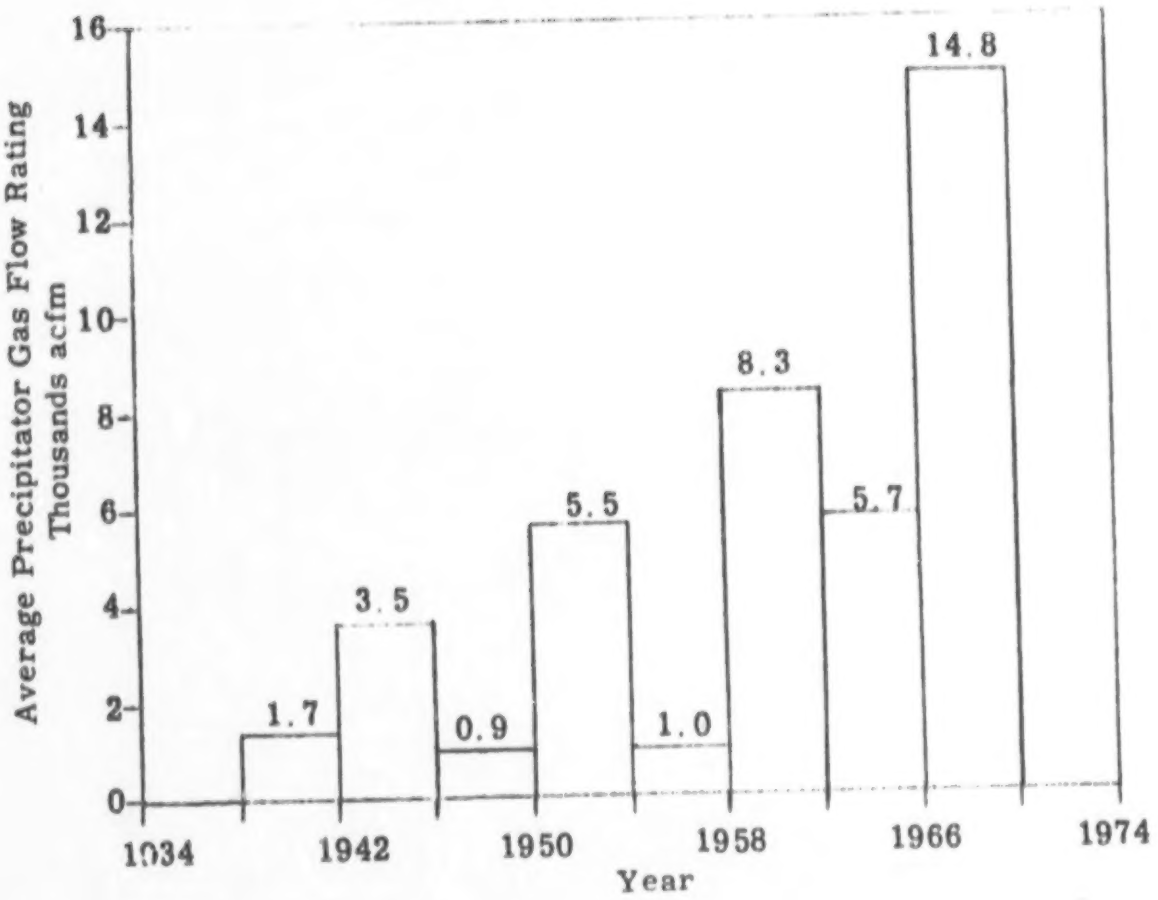

Figure 19.17. Average Gas Flow Ratings of Elemental Phosphorus Process Precipitators, 1938-1969.

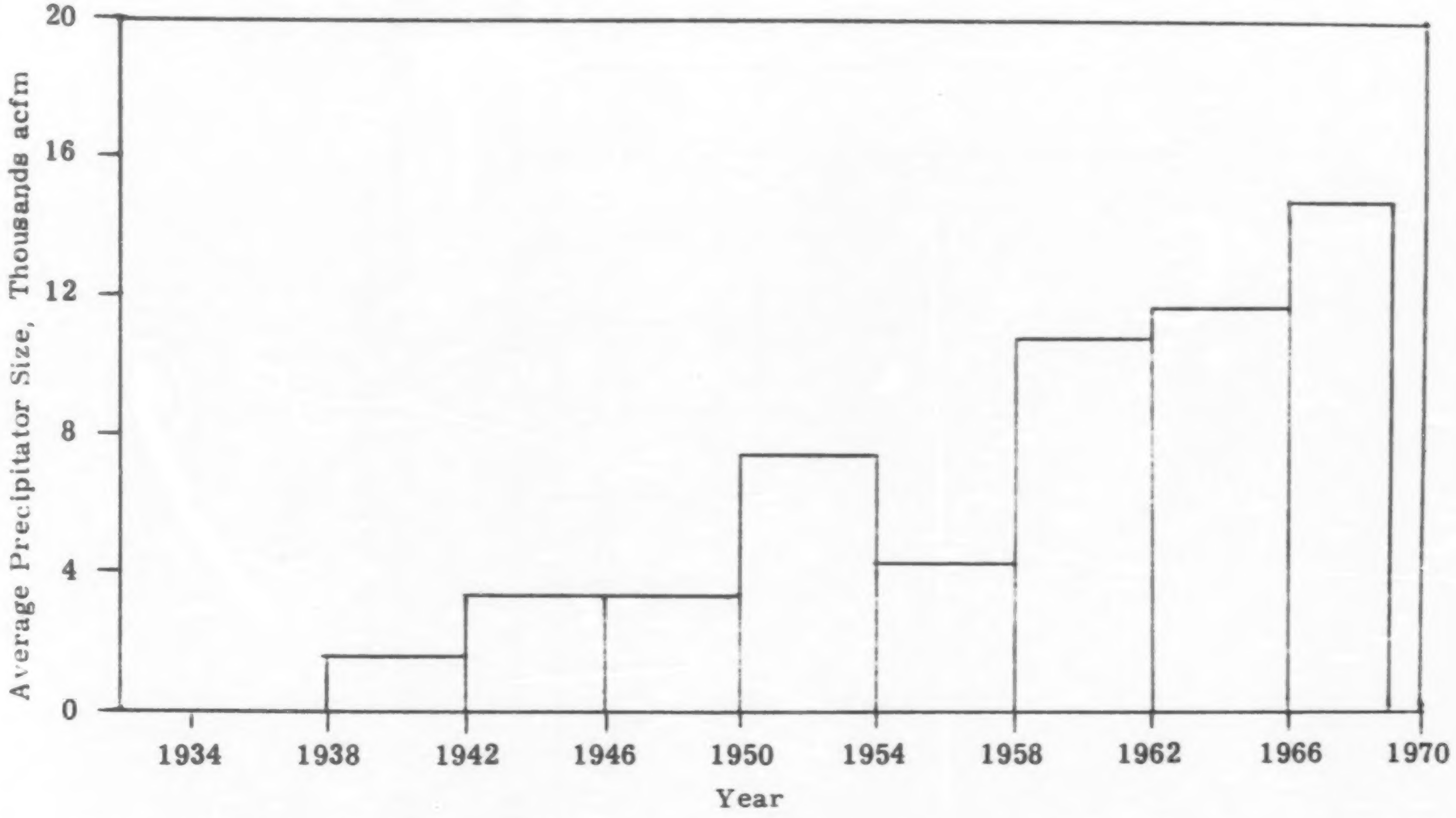

Figure 19.18. Average Size of Elemental Phosphorus Process Precipitators Over Period 1938 to 1969.

Table 19.6

Design Efficiency Trends for Hot Phosphorus Precipitators (1938-1969)

Pptr. Contract Year	No. of Install.	Pptr. Capacity acfm	Design Efficiency % Weighted on acfm Basis Annual	4-year Avg.
1938	1	260	98	
1939	2	2,650	99.9	
1940	0	0	--	
1941	1	4,000	99.9	99.8
1942	1	5,000	99.9	
1943	0	0	--	
1944	2	6,500	99.9	
1945	1	2,330	97.0	99.4
1946	0	0	--	
1947	0	0	--	
1948	0	0	--	
1949	1	3,500	99.0	99.0
1950	0	0	--	
1951	0	0	--	
1952	2	13,300	99.9	
1953	1	9,000	99.9	99.9
1954	0	0	--	
1955	1	4,300	99.9	
1956	0	0	--	
1957	0	0	--	99.9
1958	1	12,200	96.0	
1959	0	0	--	
1960	0	0	--	
1961	2	20,200	99	97.9
1962	0	0	--	
1963	1	11,400	98.3	
1964	1	12,000	99	
1965	0	0	--	98.7
1966	2	32,050	99	
1967	2	26,600	99	
1968	0	0	--	
1969	0	0	--	99.0

Table 19.7

Summary of Design Data
Electric Furnace Phosphorus Precipitator
(1938-1969)

(a) Precipitator Gas Velocity

Pptr. Gas Velocity ft./sec.	No. of Install.	Pptr. Capacity Thousand acfm In Velocity Interval	% of Installations In Velocity Interval
1.0-1.4	1	0.26	5
1.5-1.9	2	5.8	10
2.0-2.4	9	38.4	45
2.5-2.9	0	0	0
3.0-3.4	1	12.0	5
3.5-3.9	0	0	0
4.0-4.4	3	56.9	15
4.5-4.9	1	10.0	5
5.0-5.4	1	11.4	5
5.5-5.9	2	25.9	10
Totals	20	150.7	100

(b) Inlet Gas Temperature

Inlet Temp. °F	No. of Install.	Pptr. Capacity Thousand acfm In Temp. Interval	% of Installations In Temp. Interval
500-549	0	0	0
550-599	11	78.5	55
600-649	3	35.7	15
650-699	4	25.8	20
700-749	1	2.3	5
750-799	1	10.2	5
800-849	0	-	0
Totals	20	152.5	100

(c) Inlet Particulate Loading

Inlet Particulate Load gr/scf	No. of Install.	Pptr. Capacity Thousand acfm In Loading Interval	% of Installations In Loading Interval
4-5.9	0	0	-
6-7.9	2	25.4	14.3
8-9.9	1	9.0	7.1
10-11.9	0	0	0
12-13.9	9	30.9	64.3
14-15.9	2	12.5	14.3
Totals	14	77.8	100

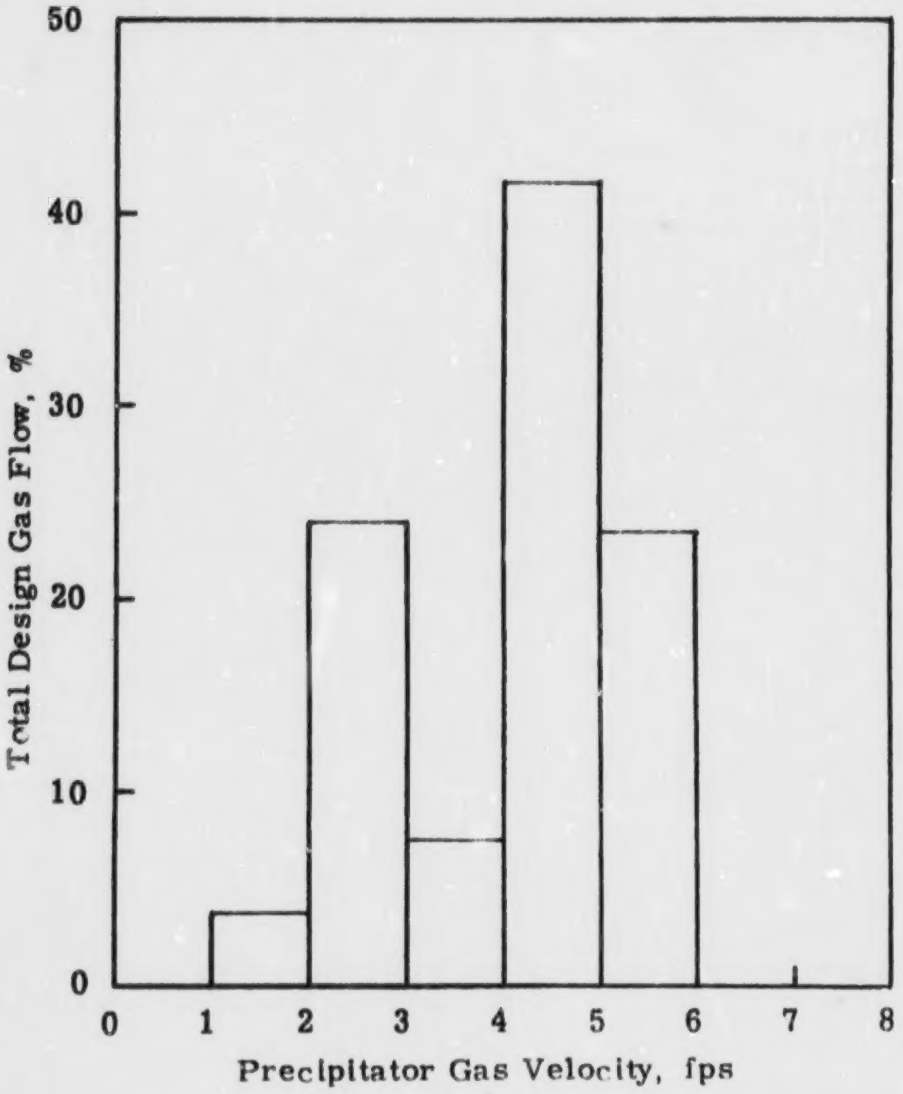

Figure 19.19. Distribution of Elemental Phosphorus Process Precipitator Design Gas Velocity, 1938-1969.

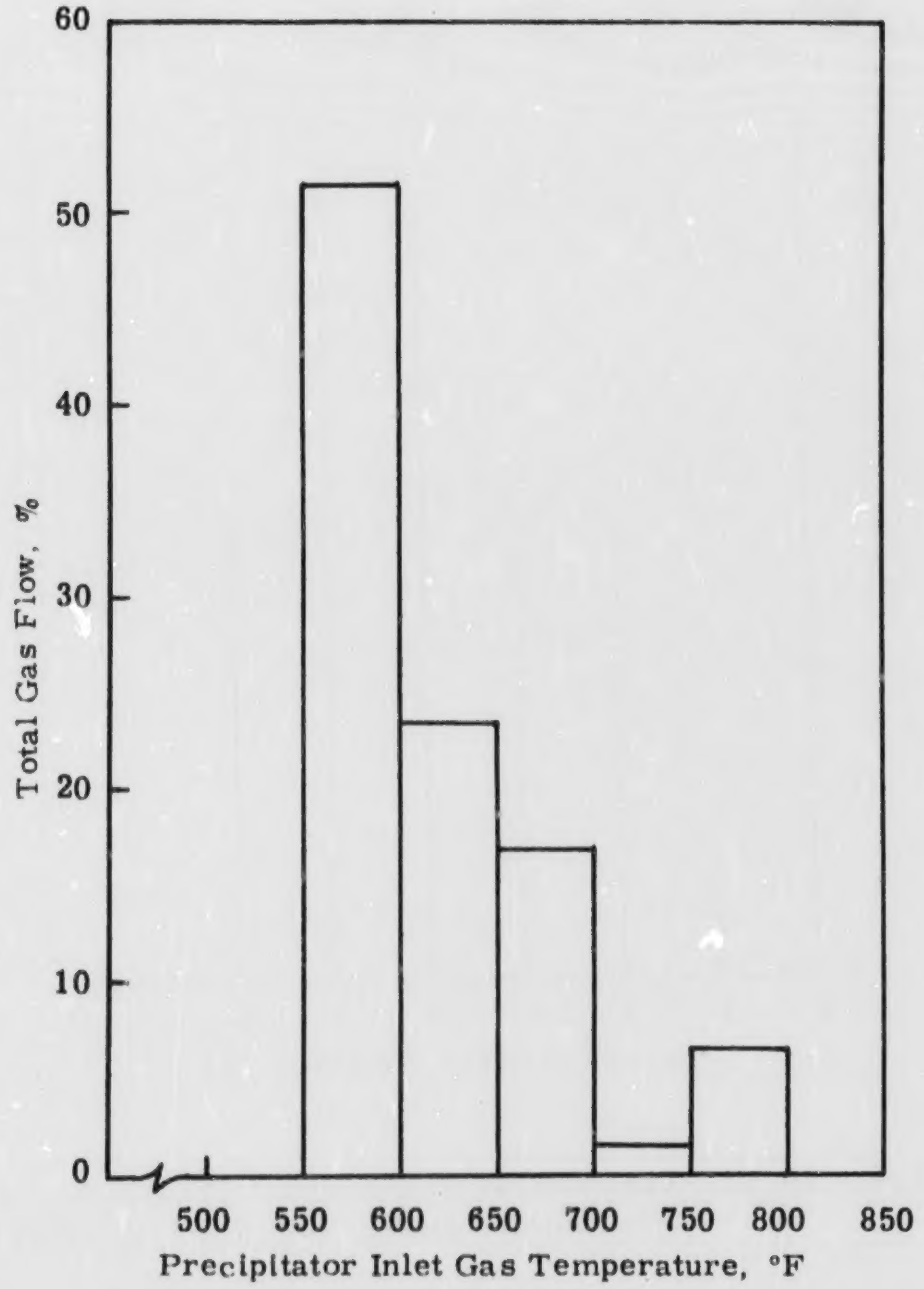

Figure 19. 20. Distribution of Elemental Phosphorus Process Precipitator Design Inlet Gas Temperature, 1938-1969.

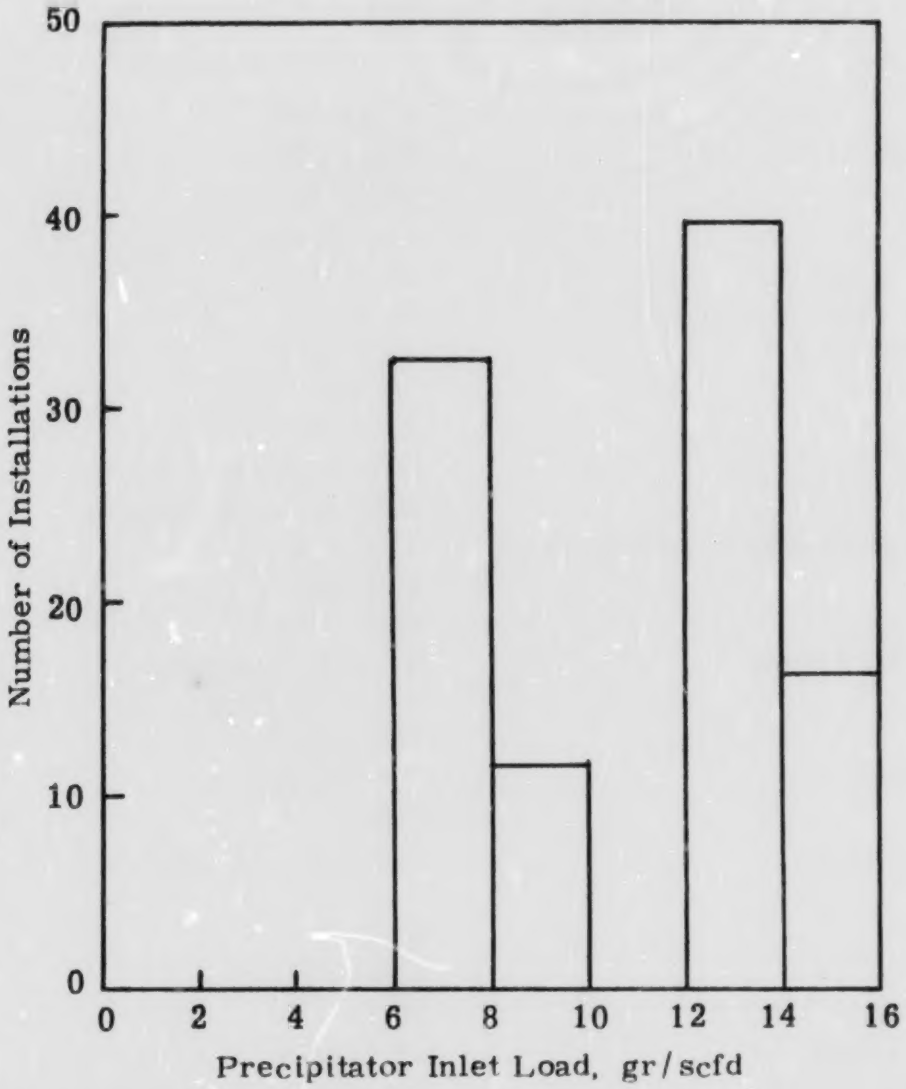

Figure 19.21. Distribution of Elemental Phosphorus Precipitator Design Inlet Loading, 1938-1969.

Table 19.8

Summary of Design Data on Two Precipitators for Cleaning Phosphorus Furnace Gas

Gas flow	3500 cfm	11,500 cfm
Temperature	600-800°F	800°F
Pressure	+0.1" w. g.	+0.1" w. g.
Gas velocity	1.9 ft/sec	4.6 ft/sec
Treatment time	16.2 sec	6.6 sec
Density of dust	30 $\mathrm{lb/ft^3}$	30 $\mathrm{lb/ft^3}$
Inlet dust loading	3 gr/acf	2 gr/acf
Inlet dust size	50-60% $<10\mu$ 30-40% $<40\mu$	
Charge	80% phosphate rock, 15% coke, 5% misc.	
Power to electrostatic precipitator	7,515 W	40 kW
Current	250 mA	500 mA
Voltage	350 V ac	
Collection electrode	96 10¾" dia tubes 15'8" long	160 11¼" dia 15' long
Spark rate	100/min	100/min
Pressure drop	0.2" w. g.	0.3" w. g.
Efficiency	99%	98%
Type power supply	full wave (2 fields)	full wave (2 fields)
DC voltage	67 kV	67 kV
DC current	250 mA/field	280 mA/field
Total treatment length	30 ft	30 ft
Collector rappers	24 shafts	8 shafts
Rapper type	Impulse, elec.-mesh	
Discharge electrode	0.25" sq wire	barbed wire
Number of wires	96	160
Total discharge wire length	1720 ft	2720 ft
Discharge wire rappers	vibrators	
Operating problems	Wire breakage, air leakage 15 wires replaced/year, tubes reamed twice/year	

Table 19.9

Electric Furnace Phosphorus Precipitator Costs

$ /acfm

Efficiency Range	Gas Volume Range - Thousands acfm					
	0-4.9		5-9.9		10-20	
	FOB	Erected	FOB	Erected	FOB	Erected
			(a) 1959-1968			
90.0-94.9	-	-	-	-	-	-
95.0-98.9	-	-	-	-	-	14.0 (1)
99.0 +	-	-	-	-	10.1 (3)	16.7 (2)
			(b) 1949-1958			
90.0-94.9	-	-	-	-	-	-
95.0-98.9	-	-	-	-	12.4 (1)	17.7 (1)
99.0 +	27.4 (2)	34.6 (2)	22.8 (1)	31.9 (2)	-	-
			(c) 1938-1948			
90.0-94.9	-	-	-	-	-	-
95.0-98.9	-	17.7 (1)	-	-	-	-
99.0 +	27.2 (1)	36.0 (3)	14.5 (1)	-	-	-

Note: Numbers in parentheses indicate number of installations in which the cost data were averaged.

Table 19.10

Average Electric Furnace Phosphorus Precipitator Costs
(1938-1969)

Period	Average Pptr. Size acfm	Efficiency % Weighted on acfm	Precipitator Cost $ /acfm FOB	Erected
1938-1941	1,728	99.8	-	22.2
1942-1945	2,458	99.4	18.7	33.5
1946-1949	3,500	99.0	27.5	35.1
1950-1953	7,430	99.9	28.0	34.3
1954-1957	4,300	99.9	-	31.4
1958-1961	10,800	97.9	12.4	16.3
1962-1965	11,700	98.7	6.8 *	14.0
1966-1969	14,662	99.0	11.8	-

*No Electrical Equipment

excess of 583°F to avoid phosphorus condensation. Electrostatic precipitators and bag-houses utilizing high temperature fabrics are, in theory, suitable gas cleaning devices. Due to the frequency of thermal upsets of phosphorous furnaces, however, with subsequent damage to the filter fabric, electristatic precipitators are perhaps more suitable for this application.

The data on design gas velocities indicate a trend to higher velocities. Figure 19.22 shows the (acfm) weighted average design gas velocity over four-year periods from 1938 to 1969.

19.6 PRECIPITATORS FOR THE PHOSPHORIC ACID INDUSTRY

Most of the elemental phosphorus produced in the U. S. (about 75% in 1965) is burned and hydrated to make phosphoric acid. In a typical installation, liquid phosphorus and air are introduced into a cylindrical combustion chamber where the phosphorus burns according to the reaction,

$$4P + 5O_2 \rightarrow 2P_2O_5$$

The gases leaving the combustion chamber, consisting of P_2O_5, N_2, O_2, steam, and nitrogen oxides, pass into a hydration tower where P_2O_5 is hydrated to orthophosphoric acid. The hydrated acid is passed through a water-spray absorbing tower, where an acid of about 85% H_3PO_4 (62% P_2O_5) is produced according to the following reaction:

$$P_2O_5 + 3H_2O \rightarrow 2H_3PO_4.$$

Traces of arsenic present in the original phosphorus are carried over into the acid, and must be removed by treatment of the acid with H_2S if the acid is to be used in food products.

Acid mist is carried over by the waste gases from the adsorber tower, and must be removed prior to exhausting the gases. Wire-mesh and glass fiber mist eliminators, venturi scrubbers, packed towers, and electrostatic precipitators can be used to control acid mist. The salvage acid is returned to the system, and the clean gas discharged to the atmosphere. Figure 19.23 is a schematic diagram of the phosphorus combustion process.

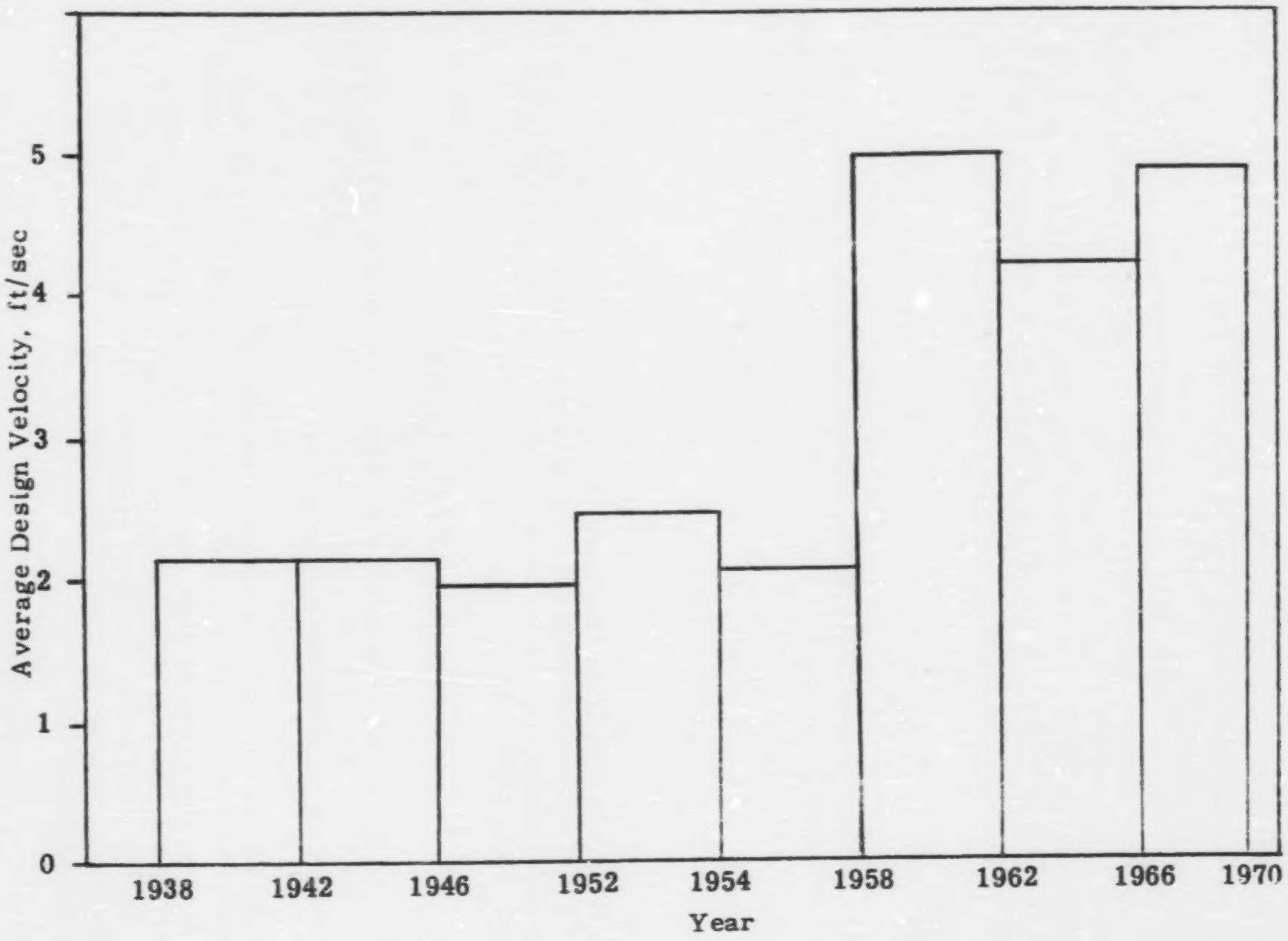

Figure 19.22. Trends in Design Velocity for Elemental Phosphorus Precipitator, 1938-1969.

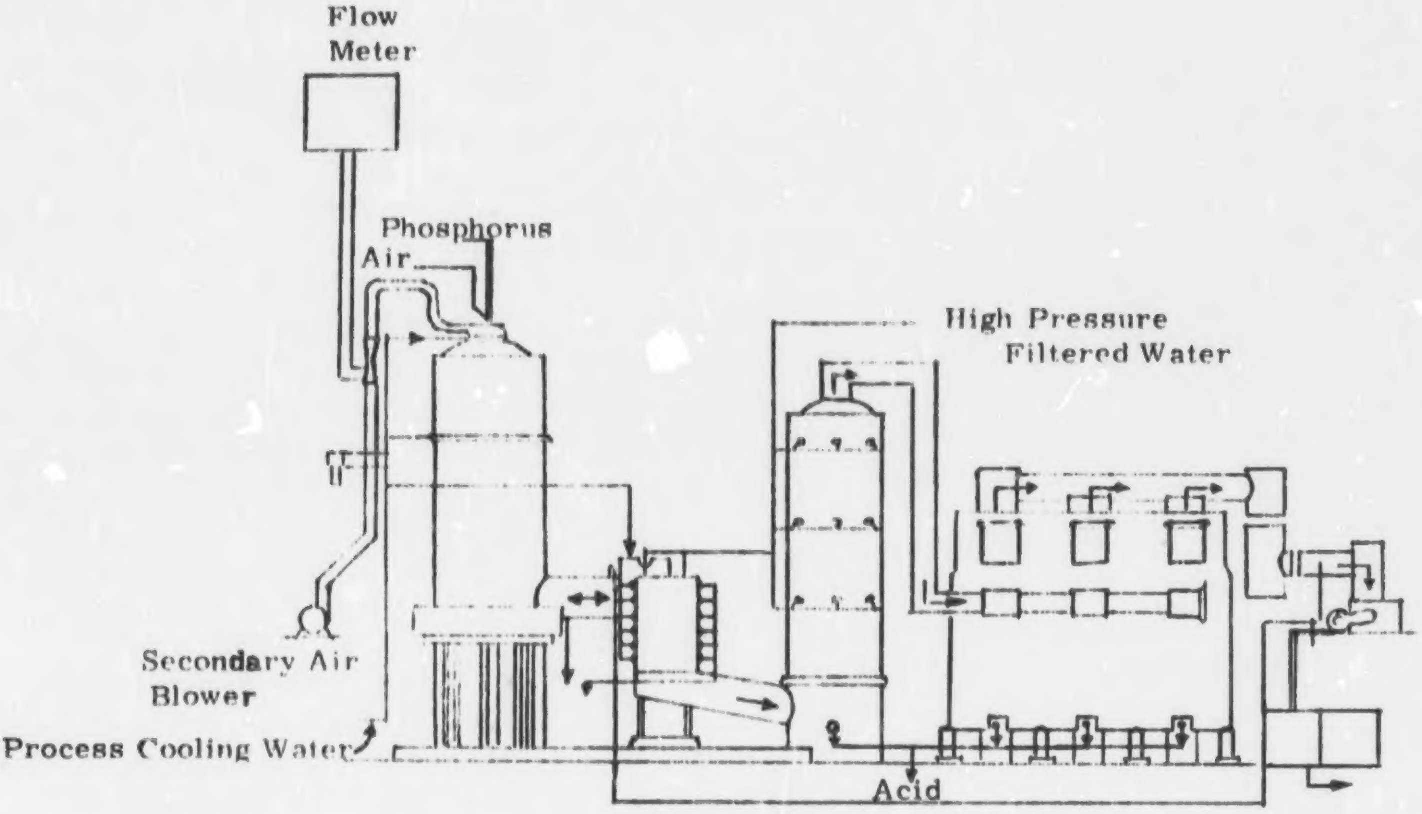
Flow Meter
Phosphorus
Air
High Pressure Filtered Water
Secondary Air Blower
Process Cooling Water
Acid

Figure 19.23. Flow Sheet for Phosphoric Acid Production by Phosphorus Combustion.

Description of electrostatic precipitator and subsystem. A typical precipitator design consists of a single-stage, vertical-flow, pipe-type unit. The shell is cylindrical in form, and is not thermally insulated. Stainless steel or equivalent construction is used. Lead is not used because the temperature may suddenly rise above the softening point of the lead.

The discharge electrodes are constructed of stainless steel and can be of various types as described in Chapter 10, Part I.

The corrosive nature of the gas creates some problems in the choice of materials and in the geometry of the support system.

Stainless steel pipes varying from 5 to 15" in diameter are employed as collecting electrodes. The arrangement used consists of a top-header only with the exterior surface of the pipe exposed to the entering gas stream. Since the precipitate is a free-flowing liquid, no cleaning of either discharge or collection electrodes is required

In the production of phosphoric acid, the various pieces of equipment are close-coupled. Because of this, the acid collected in the precipitator sump is transported from the sump, by gravity, directly to the hydrator storage. A liquid trap is used to keep the two systems isolated.

Table 19.11 is a summary of a number of phosphoric acid mist precipitator installations and precipitators, annual installed capacity, and cumulative capacity for 1927-1969. Five-year average capacity and average precipitator size over each five-year period are also shown.

Table 19.12 shows annual installed capacity, design efficiency, and weighted average design efficiency (five-year basis).

Figure 19.24 shows the cumulative installed capacity of electrostatic precipitators on acid mist phosphoric acid plants from 1928-1956.

Figures 19.25 and 19.26 show the average installed capacity averaged over five-year intervals and the average precipitator size during the same period.

Tabulations of design gas velocities and design inlet temperatures from 1927 to the present are given in Tables 19.13 and 19.14. The

Table 19.11

Phosphoric Acid Mist Precipitator Installations 1927-1956

Pptr. Contract Year	No. of Install.	Pptr. Capacity acfm	Design Efficiency % Weighted on acfm Basis	
			Annual	5-year Avg.
1927	1	3,000	--	-
1928	0	-	--	
1929	0	-	--	
1930	1	2,000	97.5	
1931	0	-	--	97.5
1932	0	-	--	
1933	1	6,000	95	
1934	1	35,800	98	
1935	0	-	--	
1936	0	-	--	97.6
1937	0	-	--	
1938	0	-	--	
1939	2	41,000	91.0	
1940	0	-	--	
1941	2	10,675	99.9	92.8
1942	0	-	--	
1943	0	-	--	
1944	0	-	--	
1945	0	-	--	
1946	0	-	--	-
1947	2	9,100	99.9	
1948	0	-	--	
1949	0	-	--	
1950	1	4,050	98	
1951	0	-	--	99.5
1952	0	-	--	
1953	0	-	--	
1954	0	-	--	
1955	0	-	--	
1956	1	26,000	85	85

Table 19.12

Phosphoric Acid Mist Precipitator Installations 1927-1956

Year	No. of Install.	No. of Pptr.	Pptr. Capacity acfm	Cumulative Capacity acfm	Average Capacity acfm	
					5-year Avg. (a)	Pptr. Basis (b)
1927	1	1	3,000	3,000		
1928	0	0		3,000		
1929	0	0		3,000		
1930	1	1	2,000	5,000		
1931	0	0		5,000	1,000	2,500
1932	0	0		5,000		
1933	1	1	6,000	11,000		
1934	1	2	35,800	46,800		
1935	0	0		46,800		
1936	0	0		46,800	8,360	13,933
1937	0	0		46,800		
1938	0	0		46,800		
1939	2	5	41,000	87,800		
1940	0	0		87,800		
1941	2	4	10,675	98,475	10,335	5,742
1942	0	0		98,475		
1943	0	0		98,475		
1944	0	0		98,475		
1945	0	0		98,475		
1946	0	0		98,475	0	-
1947	2	4	9,100	107,575		
1948	0	0		107,575		
1949	0	0		107,575		
1950	1	1	4,050	111,625		
1951	0	0		111,625	2,630	2,630
1952	0	0		111,625		
1953	0	0		111,625		
1954	0	0		111,625		
1955	0	0		111,625		
1956	1	2	26,000	137,625	5,200	13,000

(a) Pptr. capacity (acfm) divided by number of years in period.
(b) Pptr. capacity (acfm) divided by number of precipitators installed in period.

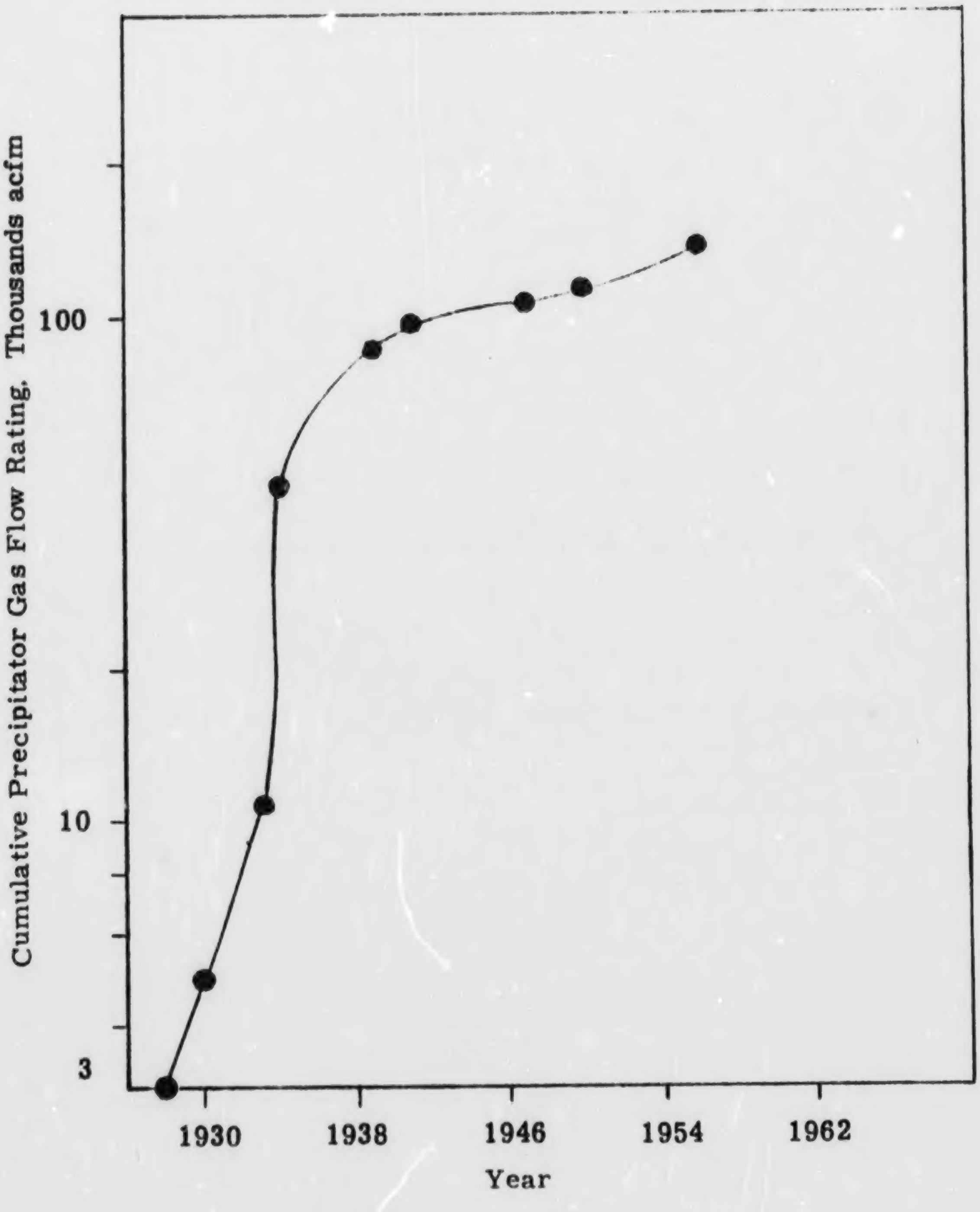

Figure 19. 24. Cumulative Phosphoric Acid Mist Precipitator Capacity Since 1928.

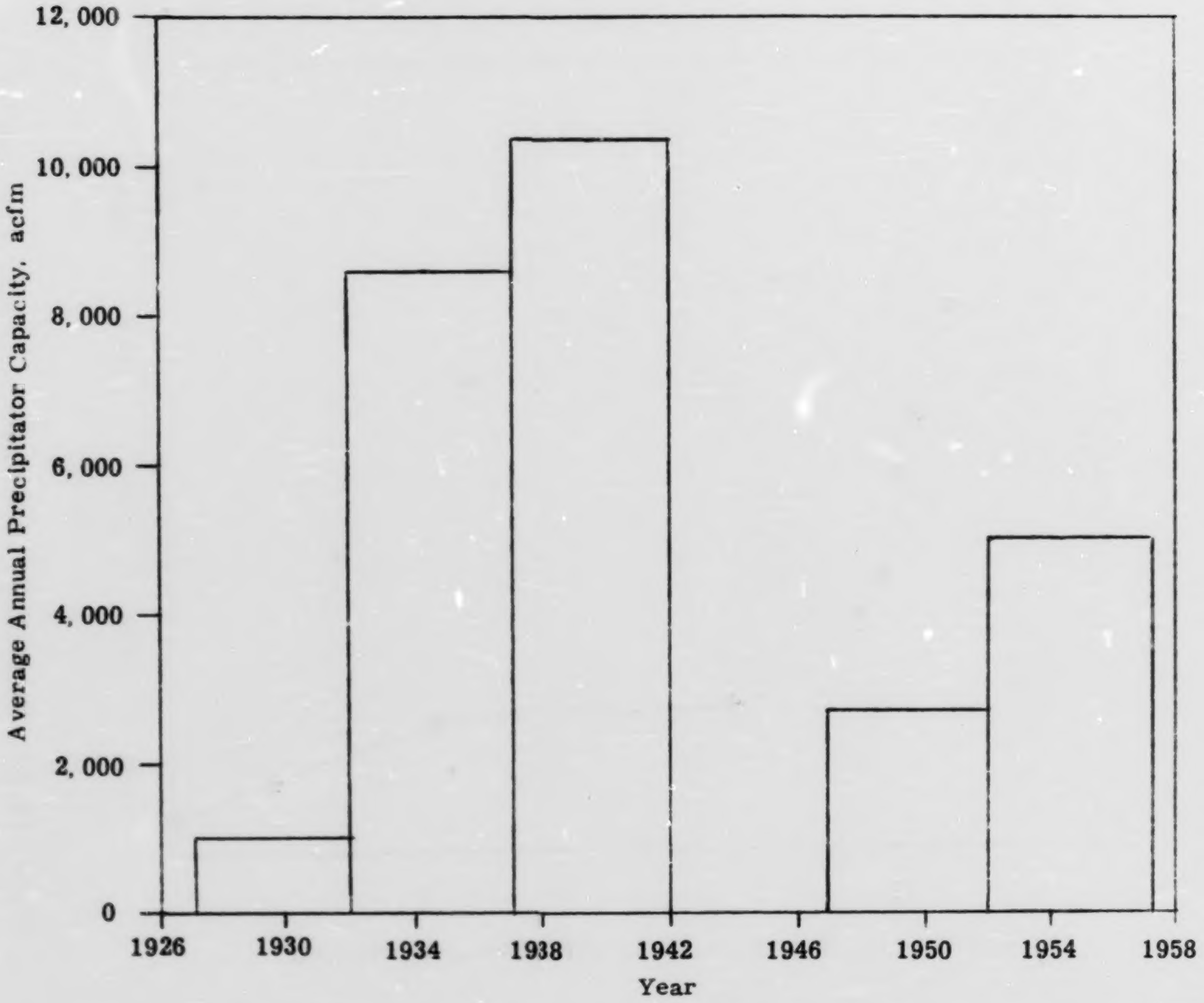

Figure 19. 25. Annual Phosphoric Acid Mist Precipitator Capacity for Period 1927-1956.

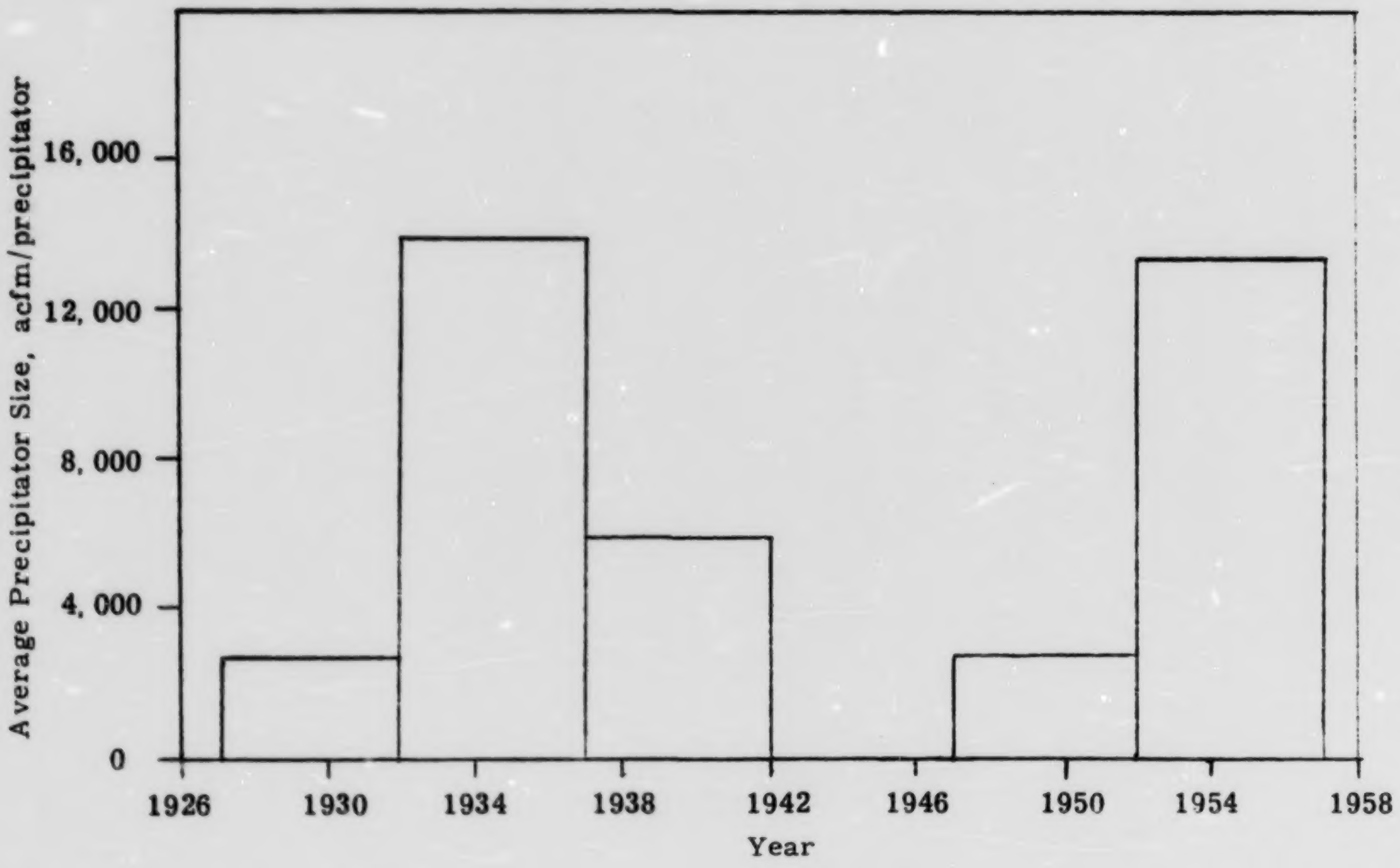

Figure 19.26. Average Phosphoric Acid Mist Precipitator Size Over Period 1927-1956.

Table 19.13

Precipitator Gas Velocity (Design)
Phosphoric Acid Mist Precipitator 1927-1969

Velocity Interval fps	No. of Install.	No. of Pptrs.	Pptr. Capacity Thousand acfm in Interval	% of Installations in Velocity Interval
2-3	1	1	1.6	8.3
3-4	4	5	22.3	33.3
4-5	1	1	3.0	8.3
5-6	4	5	48.8	33.3
6-7	0	0	0	-
7-8	2	6	62.0	16.7
Totals	12	18	137.7	100

Table 19.14

Precipitator Inlet Temperature (Design)
Phosphoric Acid Mist Precipitator 1927-1969

Pptr. Inlet Temp. °F	No. of Install.	No. of Pptrs.	Pptr. Capacity Thousand acfm in Interval	% of Installations in Interval
100-150	-	-	-	-
150-200	6	13	58.8	54.5
200-250	1	2	26	9.1
250-300	4	5	46.9	36.3
Totals	11	20	131.7	100

number of installations, precipitators, and gas volumes are also shown.

Table 19.15 is a summary of several inlet loadings.

Bar graphs of the data showing input parameters are shown in Figures 19.27 to 19.29.

Table 19.16 presents a summary of the range of actual performance data--gas velocity, inlet gas temperature, and loading--with the mean value of each parameter. Values of field strength and input power are not available.

Economics--phosphoric acid mist precipitator costs. Cost data for several precipitators have been reported in Table 19.17 for 10 to 12 year periods. Both FOB and installed costs are included when available. The limited data preclude graphical presentation. No maintenance or operating costs are available from this source.

General observations and discussion of trends - phosphoric acid mist precipitator. The data presented indicate that the use of electrostatic precipitators for phosphoric acid mist control applications has declined in recent years. Since the demand for high purity phosphoric acid from the combustion process has not declined, it is evident that other factors are responsible.

It is believed that the following are among the more important of these factors:

> Many of the precipitators were installed with mechanical rectifiers, and maintenance problems were appreciable. Further, while stainless steels are relatively resistant to phosphoric acid, maintenance problems associated with sparking and resulting corrosion have been troublesome. Lead, while resistant, lacks the useful temperature range necessary for this application in which temperature upsets are not infrequent.

The trend now appears to be toward high energy variable orifice scrubbers for this application.

Table 19.15

Precipitator Inlet Loading (gr/scf) (Performance)
Phosphoric Acid Mist Precipitator 1927-1969

Inlet Loading gr/scfd	No. of Install.	No. of Pptrs.	Pptr. Capacity Thousand acfm	% of Installations in Interval
4-10	1	2	26.0	20
10-16	1	2	35.8	20
16-22	1	2	1.6	20
22-28	0	-	-	-
28-34	2	3	13.2	40
Totals	5	9	76.6	100

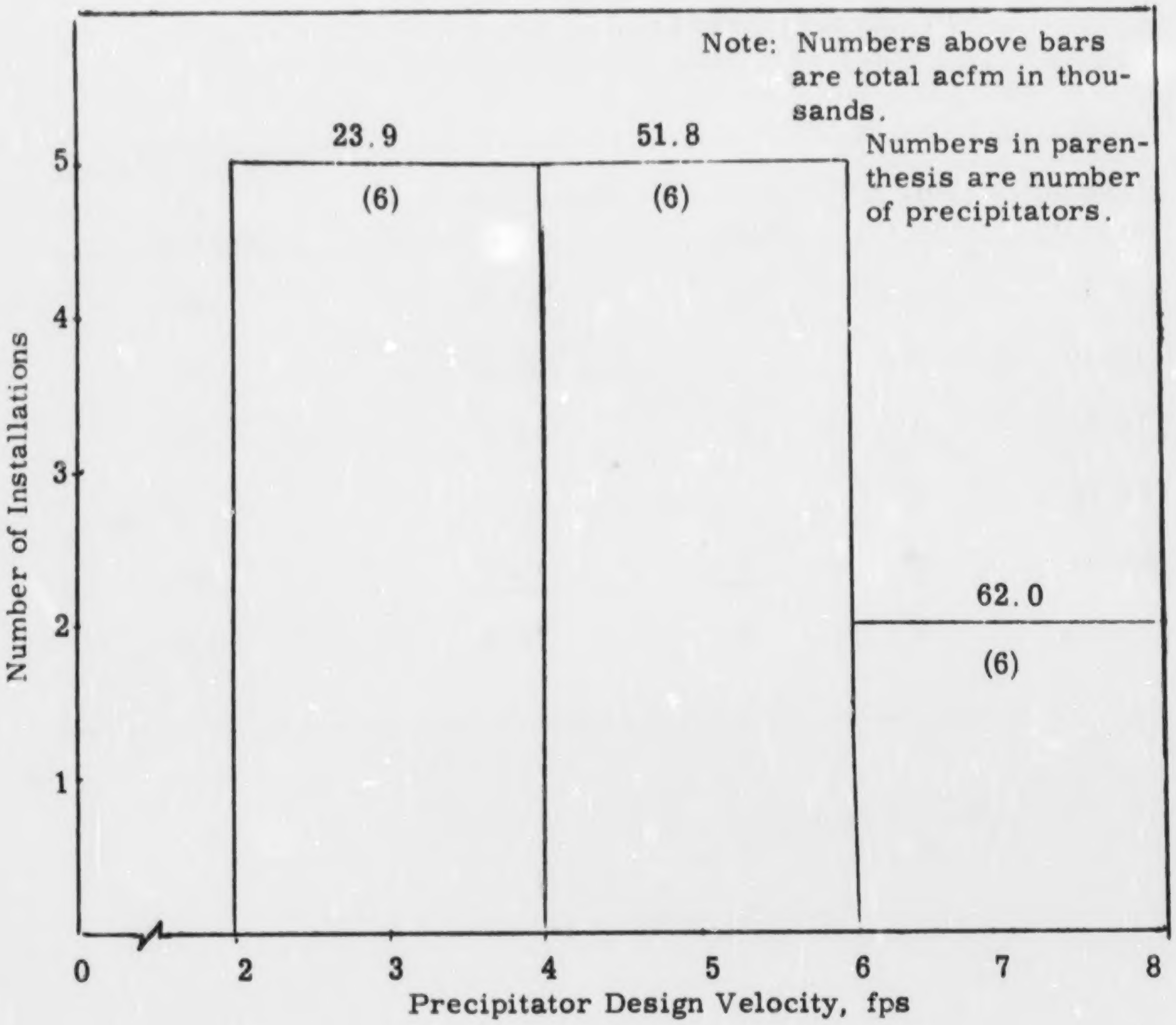

Figure 19.27. Distribution of Phosphoric Acid Mist Precipitator Design Velocity, 1927-1969.

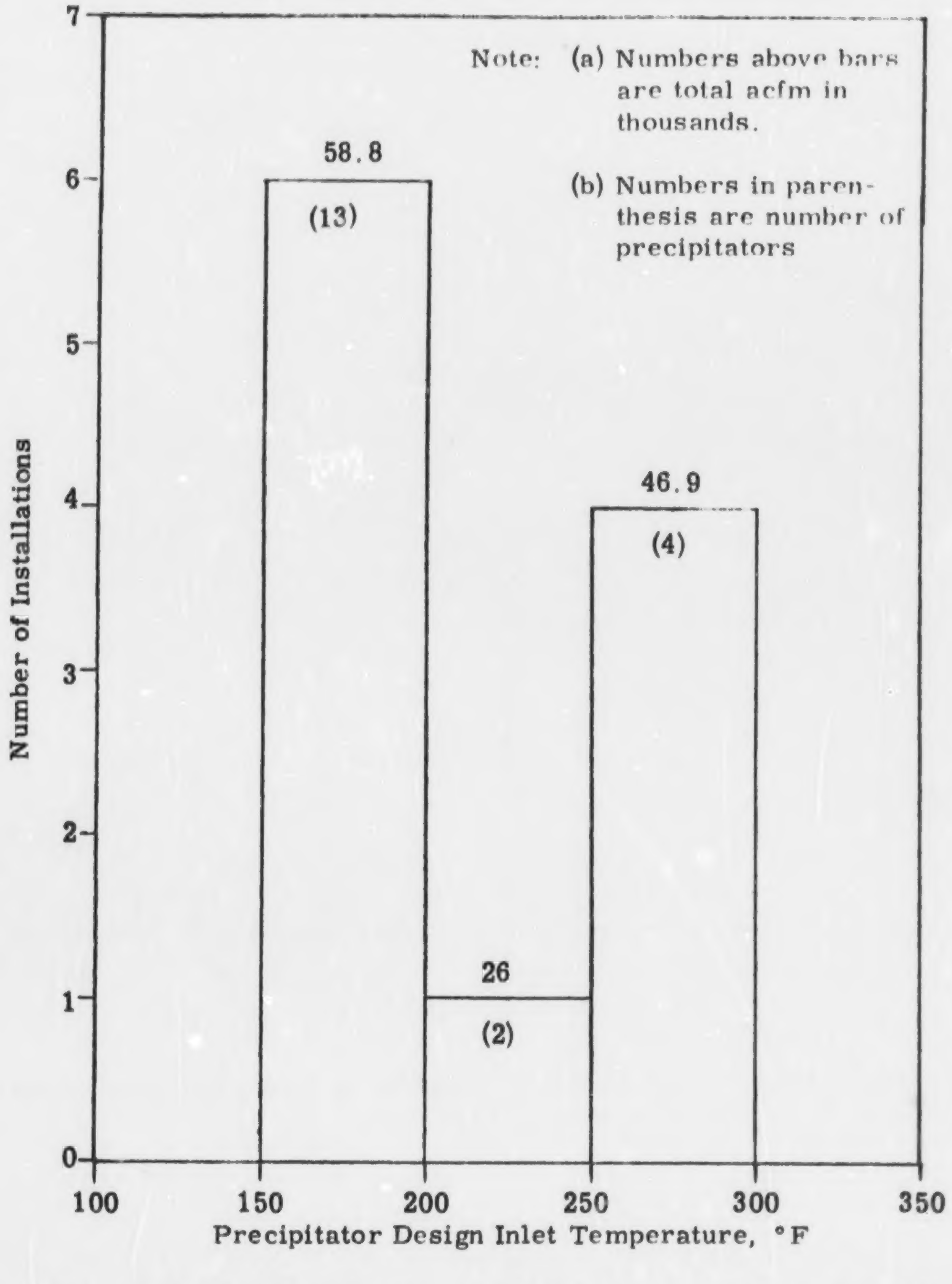

Figure 19.28. Distribution of Phosphoric Acid Mist Precipitator Design Temperature, 1927-1969.

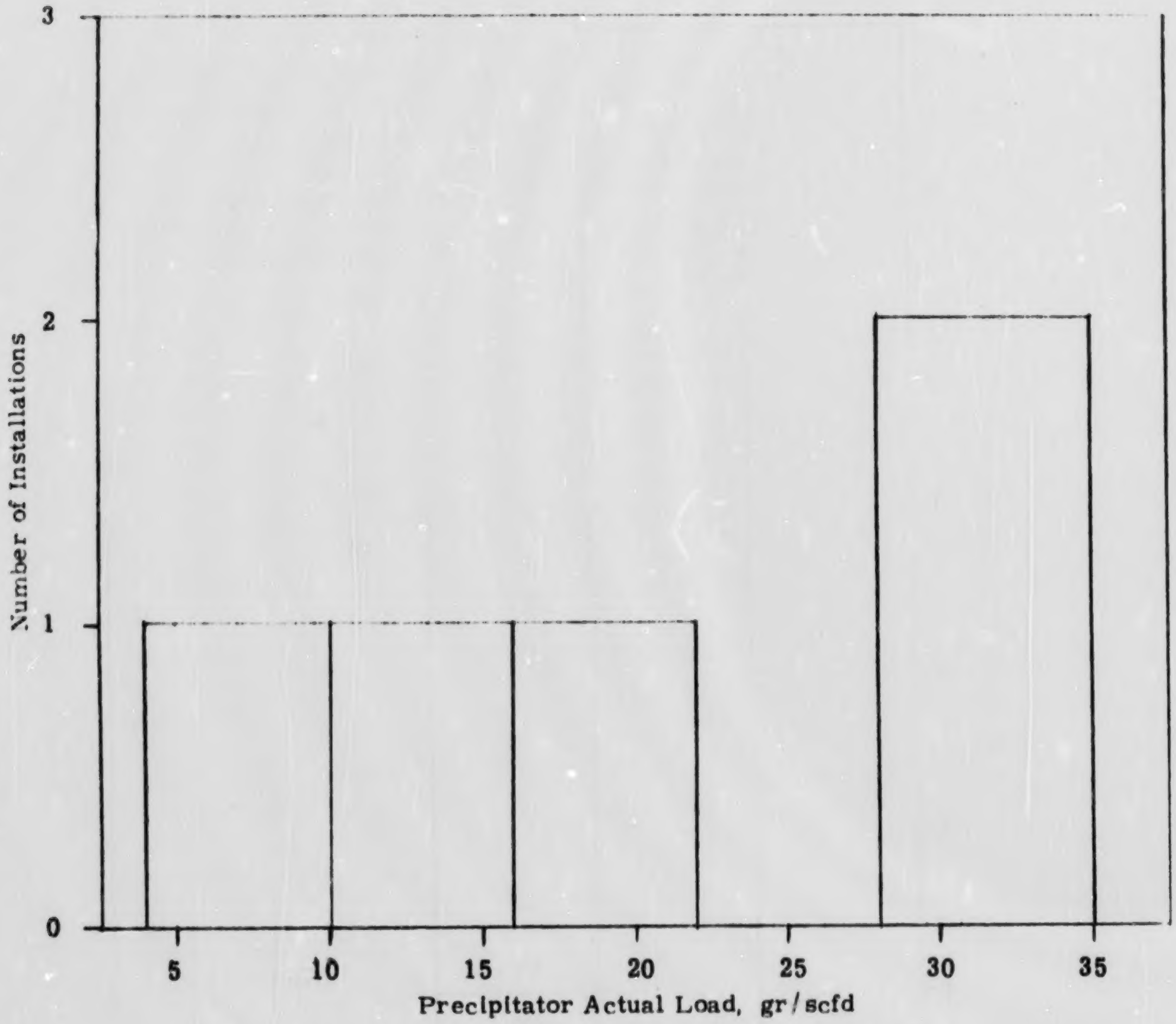

Figure 19.29. Distribution of Phosphoric Acid Mist Precipitator Actual Inlet Loading, 1927-1969.

Table 19.16

Range of Performance Data
Phosphoric Acid Mist Precipitator

Parameter	Parameter Range	No. of Install.	Mean Value of Parameter
Gas Velocity, fps	2.1 to 7.2	4	5.1
Inlet Temperature, °F	140 to 285	5	203
Inlet Loading, gr/scfd	4 to 33	5	20.3

Table 19.17

Phosphoric Acid Mist Precipitator Cost

$ /acfm

Design Efficiency Range	Gas Volume Range - Thousands acfm					
	0-9.9		10-19.9		20-40	
	FOB	Installed	FOB	Installed	FOB	Installed
			(a) 1927-1939			
Less than 95	-	-	-	-	-	-
95-98.9	-	12.4 (1)	-	-	4.90 (2)	-
More than 99 +	-	-	-	-	-	-
			(b) 1940-1950			
Less than 94.9	-	-	-	-	-	8 (1)
95-99	-	12.2 (1)	-	-	-	-
More than 99	-	19.4 (2)	-	-	-	-

Note: Numbers in parentheses are number of installations in which costs were averaged.

19.7 PRECIPITATORS FOR THE CARBON BLACK INDUSTRY

Carbon black is the ultra-fine carbonaceous product of the incomplete combustion of hydrocarbon oils and gases. The bulk of the product is used for reinforcement of rubber; speciality grades are produced for ink and paint applications.

Three major processes have been used for the production of carbon black: the furnace process, the channel process, and the thermal process. The greatest percentage is made by the furnace process, which is illustrated in Figure 19.30. Incomplete combustion is effected in a refractory lined furnace. A variety of oil and gas furnace designs is available. The particle size produced in the furnace process ($0.02 - 0.2\mu$) is intermediate between that of the channel and the thermal process.

The channel process uses an array of small natural gas flames operated with insufficient air. The black formed is deposited by impingement on the underside of moving steel channels, and is subsequently removed by stationary scraper blades.

In the thermal process, natural gas is pyrolyzed into hydrogen and carbon over heated checkerwork. The process is cyclic, with alternate heating and cracking operations. Initially, the combustion of natural gas or other fuels with air heats the furnace to about 3000° F. Following the heating cycle, natural gas alone is fed to the hot furnace where it is decomposed, producing carbon black. This process produces a rather large particle size product.

Most of the following discussion is limited to recovery from the furnace process.

The gas from the furnace (2200-2600° F) is cooled by radiation and water sprays to 450-500°F, after which fine carbon black particles are agglomerated in an electrostatic precipitator. The agglomerated particles, 0.4 to 10μ in size, are then collected in a mechanical cyclone. The product is then pelletized and packaged. About 18% of the black may be collected in the precipitator and 72% more in the cyclones. Bag filters are sometimes used following the cyclones, achieving over 99% recovery.

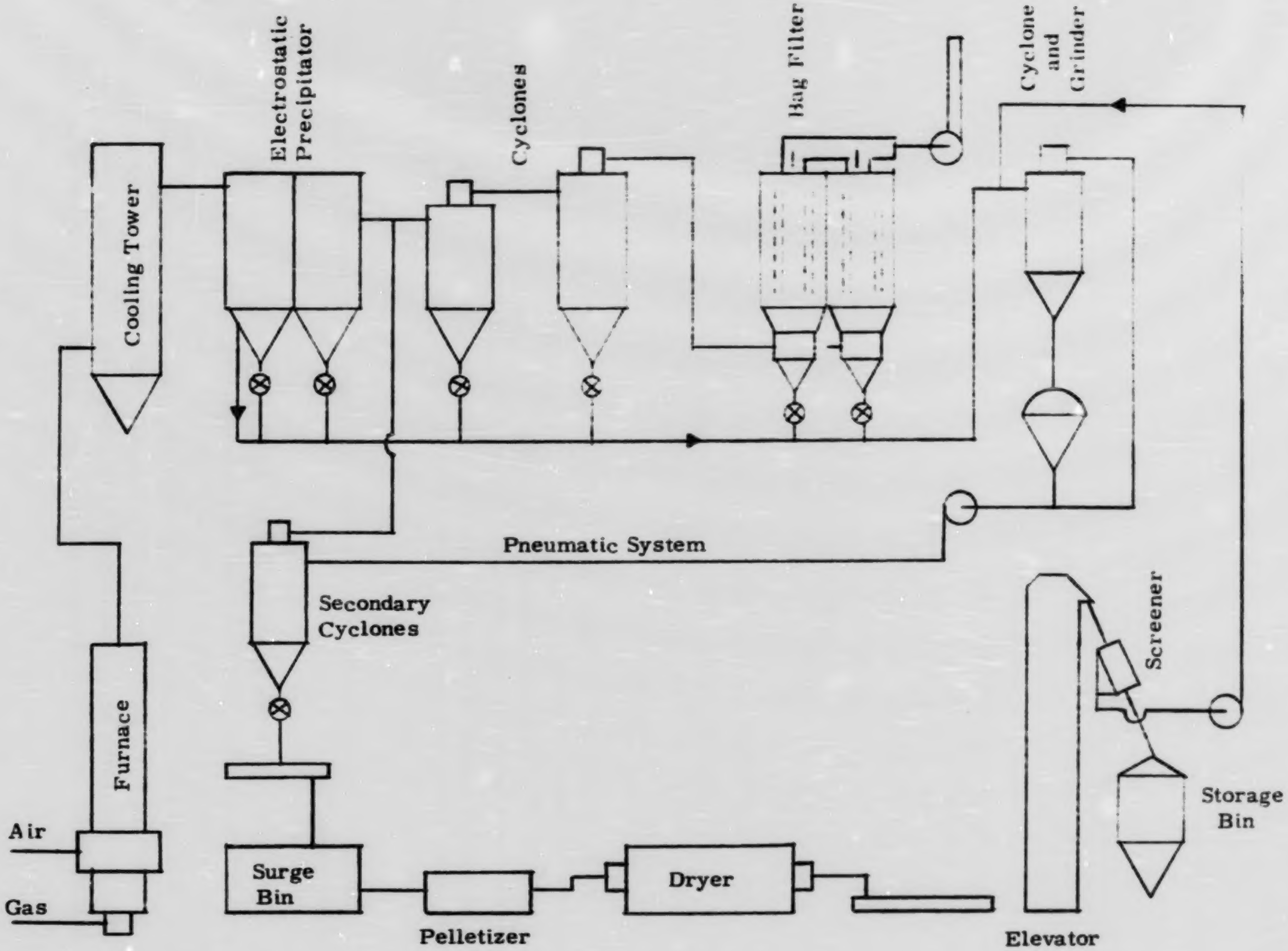
Electrostatic Precipitator
Cyclones
Bag Filter
Cyclone and Grinder
Cooling Tower
Pneumatic System
Secondary Cyclones
Screener
Furnace
Air
Gas
Storage Bin
Surge Bin
Dryer
Pelletizer
Elevator

Figure 19.30. Flow Diagram of Oil Furnace Process for Carbon Black Showing Precipitator Use.

Particulate loading varies with the process and fuel. Values are between 10 to 50 gr/scf; lower values occur with natural gas feed, higher with oil cracking processes. Average particle size in the precipitator inlet gas ranges from 0.02 to 0.4μ.

Gas composition. Typical composition of the precipitator inlet gas is as follows:

	Volume % (Wet Basis)
Carbon Dioxide	3.0
Oxygen	0.3
Carbon Monoxide	6.0
Hydrogen	8.0
Acetylene	0.5
Methane, Ethane, etc.	0.2
Nitrogen	41
Water	40

Description of electrostatic precipitator and subsystem. The precipitator most commonly used in the dry carbon black process is a single-stage horizontal flow duct type precipitator. The shell is heat-insulated and provided with a preheat system, purge vents, and explosion hatches. Both inlet and outlet flues are provided with isolation dampers so that the precipitator-cyclone system can be completely shut off from the rest of the plant during heatup and purging.

Electrode system. In general, the high voltage discharge electrode and support systems used are those described in Chapter 10, Part I. Because of the relatively high corrosion rate, a form of rod curtain collecting electrode may be used.

Collected carbon black sticks to both the discharge as well as the collecting electrodes very tenaciously, thus requiring strong cleaning methods. Both discharge electrodes and collecting electrodes are usually cleaned by the use of air vibrators.

As the precipitator is used chiefly as a particle agglomerator with the major fraction of the collection subsequently effected in a cyclone, or baghouse, no specially designed gas flow control devices are used.

Precipitate removal systems. Precipitate removal from the bottom of a carbon black collector is by either a combination scraper-screw conveyor from a flat bottom pan, or by a screw conveyor in a trough-type hopper. In both cases, the black is conveyed directly to a pelletizer.

Carbon black precipitator applications. Very little information is available on carbon black precipitators. It is estimated that nearly one hundred sets of combined precipitator-cyclone units have been built since 1926. Some limited and incomplete data for the period 1942 to 1956 are presented in Table 19. 18. No domestic carbon black installations have been noted since 1958. It should be noted that design efficiency values are based on overall precipitator-cyclone performance. No breakdown of individual performance value is available for the precipitator or cyclone.

Design efficiency values were not reported for all installations. Moreover, in some cases, more than one efficiency value is specified; this refers to the performance for different grades of black which were collected in the same installation.

Table 19. 18

Carbon Black Precipitator Installations 1942-1956

Year	No. of Install.	No. of Pptrs.	Capacity acfm	Design Efficiency %
1942	2	6	240,000	90
1943	5	10	123,950	~92
1944	1	1	40,000	90
1945	1	4	80,000	90
1947	1	1	40,000	96
1948	1	1	75,000	96
1952	2	4	120,000	93/97
1956	1	1	40,000	90

Cost Figures are quite limited and pertain to the period 1942-1952. Values reported are shown in Table 19.19.

Table 19.19

Carbon Black Precipitator
Cost Data (1942-1952)

Year	Installation Capacity	Design Efficiency	Cost $ acfm FOB	Install.	No. of Cases
1942-1945	80,000-160,000	90	---	1.50	(2)
1952	80,000	93/97	3.7	4.45	(1)

Differences in cost result from difference in design efficiency as well as auxiliaries included in some jobs.

The range of statistics reported is summarized in Table 19.20.

Table 19.20

Range of Performance Statistics
Carbon Black Precipitators (1942-1956)

Parameter	Range	Mean Value	No. of Data
Actual Gas Velocity	3.2 - 8.0 FPS	5.6	6
Inlet Gas Temperature	410 - 500°F	459	8
Inlet Particulate Load	6 - 50 gr/scf	22.3	9
Field Strength		NO DATA	
Input Power		NO DATA	

General observations, design methodology, and discussion of trends. The precipitator has served for years as a means of agglomerating carbon black particles, some as small as 0.02μ. The high conductivity of the black particles prevents good collection. Cyclones follow and remove the bulk of the black, although additional control devices are now being required for air pollution control. Considering the limited hopper catch which can be directly attributed to the precipitator, other agglomerating techniques appear to be gaining favor. In general, a variety of collection control devices is being used in series. Baghouses play an important role in many current collection schemes.

CHAPTER 19
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CHAPTER 20
THE APPLICATION OF ELECTROSTATIC PRECIPITATORS IN CLEANING MUNICIPAL INCINERATOR DUSTS

While a great many small and intermediate scale incinerators have been designed and utilized for combustion applications such as household trash and industrial waste products, the major application for electrostatic precipitators is on large municipal incinerators.

These incinerators, which are most often centrally located in an urban area, have become an increasingly important means of reducing combustible refuse to an inert residue. Due primarily to widely varying waste composition and burning characteristics, these large incinerators are a potential source of particulate emissions.

In Europe, electrostatic precipitation is widely utilized for particulate control on municipal incinerators, and a few installations are being built in the United States.

Because of the limited number of installations, statistical data for this application are sparse; however, the more significant factors and trends in the use of precipitators for this service are being discussed in this report.

20.1 TYPES OF INCINERATORS

While some municipal incinerators in the United States are designed to utilize the heat generated for such purposes as space heating, water heating, preheating combustion air, or providing steam for plant equipment, most are designed solely to reduce refuse to an inert residue. In Europe, heat utilization from incineration is practiced extensively for electric power generation and steam heating.

There are two principal furnace designs in general use for municipal incinerators--batch and continuous feed. A 1966 survey of U. S. incinerators shows that batch feed furnaces are generally limited to a maximum capacity of about 250 tons per day. Larger furnaces are the continuous feed type, and the trend in recent years has been to this type.

Incinerator furnace walls may be either water cooled or refractory lined. Water-cooled walls achieve longer furnace life, and are capable of withstanding higher temperatures.

Many municipalities operate their incinerators only five days a week or on the days of refuse collection. This presents loading, startup, and shutdown problems. Where the heat content of the refuse is low, or where the moisture content is high, incinerators may require an auxiliary fuel. In some cases, auxiliary fuel is used only on startup.

20.2 INCINERATOR CAPACITIES

Refuse incinerators are made in a wide variety of sizes to service apartment buildings, office buildings, and commercial establishments, however, the principal use of electrostatic precipitators has been on large municipal incinerators. The following discussion will be limited to these applications.

A comprehehsive survey made in the United States in 1966 shows a trend toward increasing furnace and plant capacities. These data are shown in Table 20.1.[1] The largest multifurnace plants reported in this survey process 1200 tons of refuse per 24-hour day, and the largest European plant reported a total capacity of 2400 tons of refuse per day.

Table 20.1

Range of Plant and Furnace Capacities for Municipal Incinerators Installed and Rebuilt from 1945-1965 - U. S. and Canada

Interval	Reported No. in Interval	Plant Capacity tons/day		Furnace Capacity tons/day	
		Low	High	Low	High
1945-49	16	50	600	50	175
1950-54	44	80	1000	50	250
1955-59	52	50	1200	50	300
1960-65	60	100	1200	50	350

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[1] Refer to the bibliography for this chapter.

20.3 REFUSE PROPERTIES

Refuse consumed in municipal incinerators includes varying amounts of rubbish, trash, garbage, and other wastes. The composition varies, depending upon the season, weather, location, standard of living, etc. Furnace ash can make up more than 60% by weight of refuse in winter where coal is the predominant residential fuel. In the United Kingdom, Belgium, and Czechoslovakia, ash forms a large percentage of refuse, comprising 30-65% of the total. Paper constitutes a large portion of the refuse in the United States, Canada, and the Scandinavian countries, ranging from around 40 to 70% of the total, but is as low as 3% in Poland. The metal and plastic content of refuse is greater in the United States than elsewhere. High moisture content garbage forms a greater percentage of Japanese refuse than it does in North America or Europe.[2,3]

Bulky refuse is composed of such things as logs, tree stumps, truck tires, furniture, and mattresses. This type refuse is usually collected separately and burned in special incinerators. Certain industrial wastes such as oil sludges, tars, polymers, and rubber chunks are sometimes given special treatment.[4]

The composition of the refuse from a group of selected countries is given in Table 20.2. These countries were selected because most published literature on incinerators concerns these countries.[2,3]

Table 20.2

Refuse Composition of Selected Countries (weight percentage)

	Ash	Paper	Organic Matter	Metals	Glass	Misc.
Canada	5	70	10	5	5	5
France	24	30	24	4	4	14
Japan (Tokyo)	19	25	41	3	3	9
Sweden	0	55	12	6	15	12
United Kingdom	30-40	25-30	10-15	5-8	5-8	5-10
United States	10	42	23	8	6	11
West Germany	30	19	21	5	10	15

Refuse composition affects incineration and air pollution control in several ways. Residential furnace ash has little, if any, heating value. Therefore, where ash content is high, the heating content of the refuse is low. Likewise, where refuse contains a large percentage of paper, the heating value is high. In some situations where the heating value is low, supplemental fuel, such as natural gas, oil, or coke, is used in incinerators. Generally, the heating value of refuse is higher in the United States than in Europe and Japan as shown in Table 20.3.

The moisture content of refuse varies widely, with the main variable being the garbage content. Moisture in the refuse reduces the flame temperature, which decreases the burning rate, and in some cases the refuse must be dried before entering the combustion chamber of the incinerator.[2]

Refuse in the United States contains more metals and plastic than that of Europe or Japan. Low melting point metals can cause trouble by welding to, and interfering with, the operation of moving parts of incinerator grates. When chloride-containing plastics are incinerated, corrosive hydrogen chloride fumes are formed, and incinerator components must be made to contend with this corrosion problem.[2,5]

The density of refuse, which varies with its composition and moisture content, is estimated to be from 300 to 400 lb per cubic yard in the United States.[1]

20.4 EMISSION PROPERTIES

The amount and particle size range of particulate emissions leaving an incinerator combustion chamber vary in different situations. Such factors as refuse composition, method of feeding, completeness of combustion, and operating procedures are responsible for these variations.[6]

The combustion of paper illustrates the effect of composition on particulate matter properties. Paper ash is generally of a relatively large size, has a low specific gravity, and has low electrical resistivity. These properties must be considered when selecting air pollution control equipment where the paper content of the refuse is high.

Table 20.3

Incinerator Heating Value

	Maximum	Minimum	Median
United States (41 plants reporting from 1950 to 1966)	8000 Btu/lb	3375 Btu/lb	5043 Btu/lb
Paris	4500 Btu/lb	1600 Btu/lb	
Tokyo	2345 Btu/lb	902 Btu/lb	

Reciprocating grate and rotary kiln type incinerators have a larger amount of particulate matter at the furnace exit than the other common types of plants.

The rate of furnace emission has been found to vary from 10 to 60 lb of dust per ton of refuse burned. Modern incinerators operate with an emission rate of about 35 lb per ton of refuse. Based on using 50% excess air for combustion, the 35 lb per ton can be expressed as 1.58 grains per cubic foot.[7]

The particle size range of the particulate matter also varies considerably. Usually, the same factors which decrease the amount of particulates will decrease the size of the individual particles. Figure 20.1 gives particle size distribution of fly ash measured at the point where it leaves the combustion chamber of an incinerator.[7]

The density of the dust which leaves the incinerator combustion chamber varies from about 125 lb per cubic foot to about 187 lb per cubic foot.[7]

The electrical resistivity of the particulate matter at the outlet of the combustion chamber varies with temperature and particle size distribution in addition to refuse composition, completeness of combustion, and moisture content. Figure 20.2 shows the resistivities of various samples taken from municipal incinerators in the United States and Great Britain.

Resistivity characteristics for the U. S. samples are for the less than 74-micron diameter size fraction taken from the combustion chamber. Resistivity values for the total sample were not obtained due to the high conductivity and carbon content of the coarser particles, according to the authors.[8]

Resistivities for the samples from Great Britain are shown for moisture conditions corresponding to the outlet of an incinerator (45°C water dew point), and after evaporative spray cooling (75°C water dew point).[9]

The temperature of the exhaust gases from incinerator combustion chambers is dependent upon several variables. These include type of

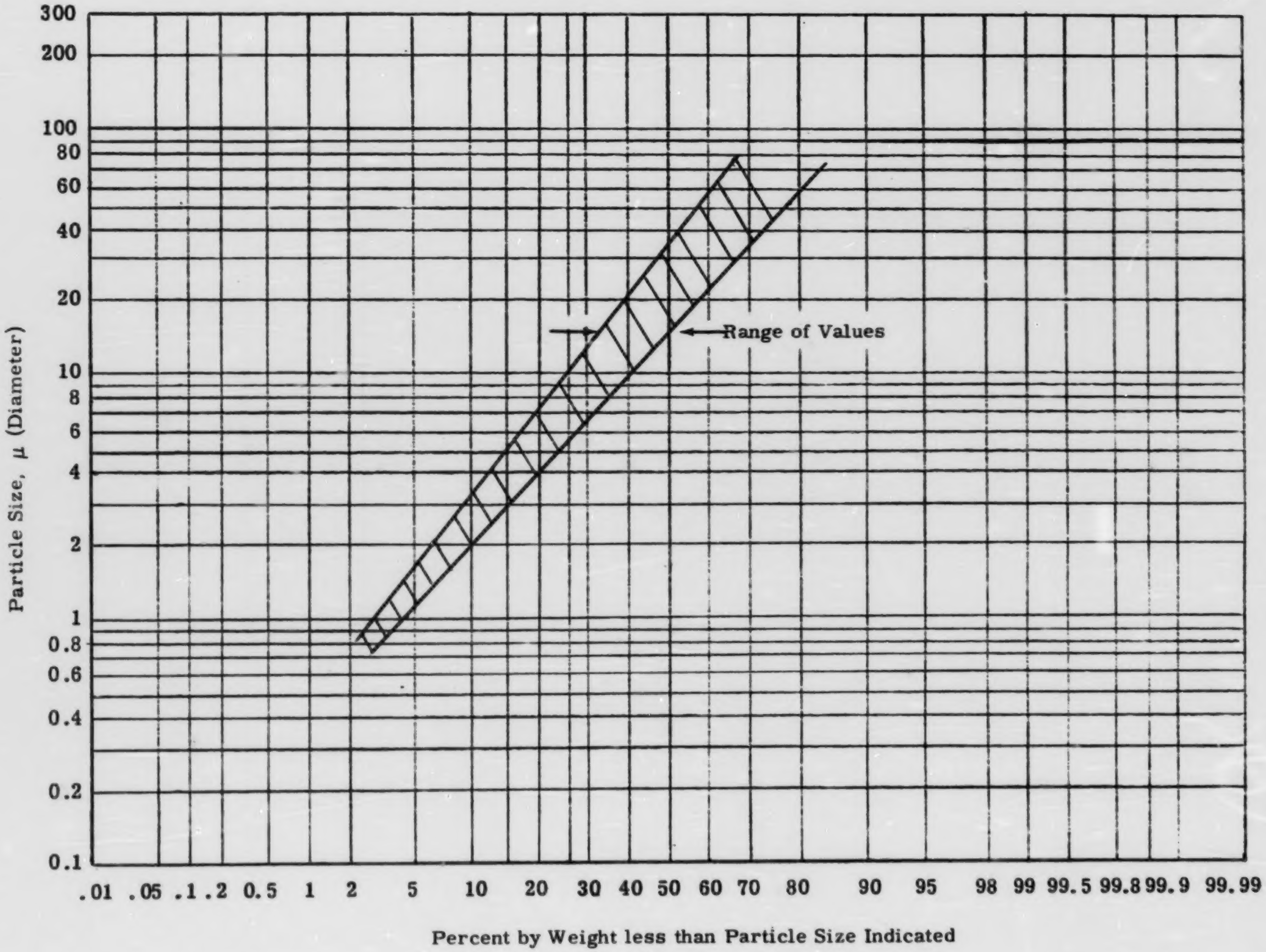
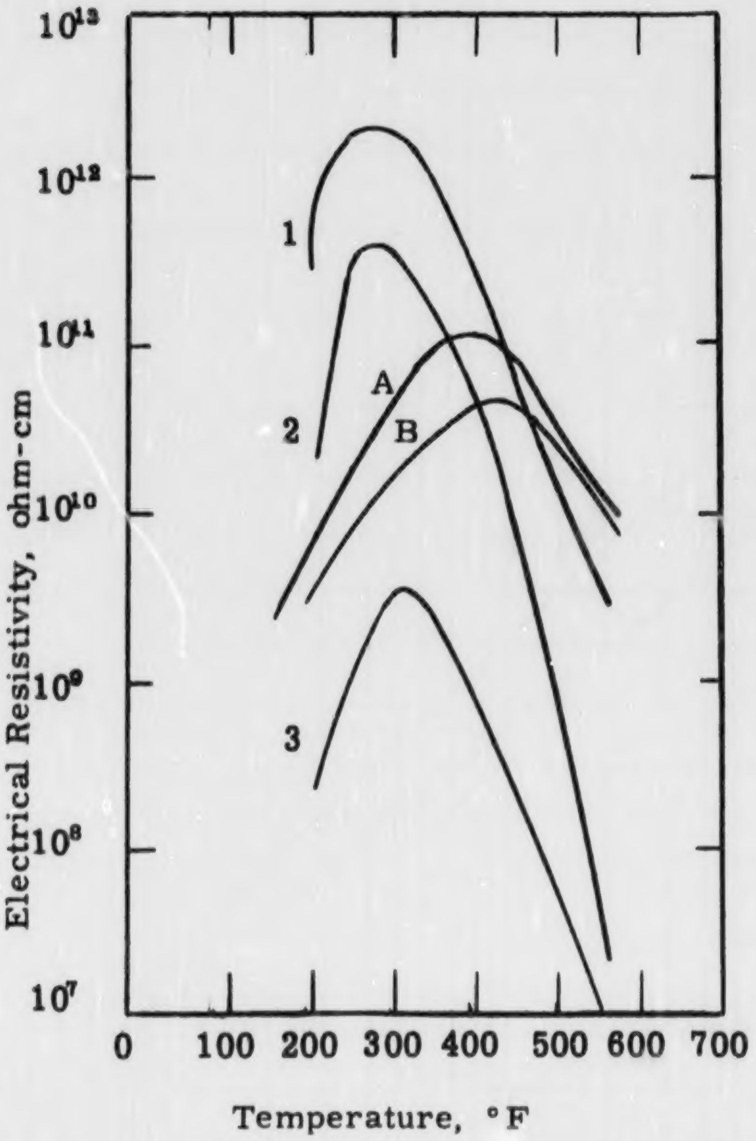


Figure 20.1. Particle Size Distribution of Incinerator Emissions Prior to Conditioning.



1-2-3: Resistivity of the less than 74μ fraction from three installations (Ref. 8).

A: Water dewpoint 45° indirect heat extraction (Ref. 9).
B: Water dewpoint 75°C evaporative spray cooling (Ref. 9).

Figure 20. 2. Electrical Resistivity of Municipal Incinerator Dust.

furnace construction, heating value and moisture content of the refuse, completeness of combustion, amount of excess air, use of auxiliary fuel, and ambient temperature. Temperatures of at least 1500°F are required for the burnout of smoke and the oxidation of most odorous compounds. Exhaust gas temperatures as high as 2500°F are reported for furnaces with water-cooled walls, but they must be held at lower levels for refractory furnaces.

Composition of the gases from incinerators varies according to the products being burned. Average and maximum values for various gases from a number of incinerators in the United States[8] and the United Kingdom[9] are listed in Tables 20.4 and 20.5, respectively.

20.5 INCINERATOR EMISSION CONTROL

One of the earliest used, and least efficient methods of dust emission control on municipal incinerators is the dry settling chamber. Combustion gases enter the chamber where the velocity is greatly reduced. This allows combustion of some of the suspended particles to be completed, and some of the larger particles are deposited. Up to 20% of the ash dust can be removed this way.[7]

In some installations, a water spray is used in combination with a settling chamber. This combination can remove up to 30% of the fly ash. The water spray is useful in reducing the temperature of the gases for downstream equipment such as emission control equipment or induced draft blowers.[7]

Mechanical collectors in which centrifugal force separates dust from the gas are frequently utilized. Collection efficiency depends on particle size, and ranges from 75 to 80% for particles larger than 20 microns. Efficiencies drop off rapidly for particles less than about 10 microns in size.

Wet scrubbers which remove dust particles by impaction on water drops can be effective in ash removal from incinerator gases. The efficiency of these scrubbers is dependent upon the relative velocity and drop size of the water. Scrubbers are highly efficient in particulate removal. The resultant humidity of the stack gas is high, so

Table 20.4

Incinerator Emission Properties—United States

	Installation No. 1		Installation No. 2		Installation No. 3	
	Maximum	Average	Maximum	Average	Maximum	Average
Temperature, °F	1673	1469	1360	1211	1714	1593
Water Vapor, % by vol	24.2	7.8	4.9	4.4	18.7	8.7
CO_2, % by vol	7.3	6.0	6.7	5.0	7.8	7.0
O_2, % by vol	15.5	13.7	14.9	13.8	14.8	13.3
Volume (thous), acfm**	170.0	143.0	158	142	72.0	64.9
Volume (thous), sdcfm*	45.2	33.8	45.2	40	14.7	14.1
Refuse Charged, tons/hour	15.8	12.9	11.0	9.8	5.3	5.1
Underfire Air, scfm/sq ft of grate	68.7	41.8	130	105	21.9	17.5
Particulate Emission, gr/sdcf*	0.745	0.553	0.820	0.694	0.540	0.380
Particulate Emission, lb/hour	231	156	308	241	56.3	46.1
Particulate Emission, lb/ton refuse charged	18.9	12.4	36.6	25.1	11.0	9.1

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* Standard dry gas at 29.92 in. Hg and 32°F.

** cfm at actual conditions of temperature and humidity.

Table 20. 5

Analysis of Gases from Incinerator Grates

		Average Reading	Maximum Reading
Oxides of nitrogen	NO & NO_2	100 ppm	200 ppm
Hydrogen sulfide	H_2S	Nil	Nil
Carbon dioxide	CO_2	6.0-7.0%	11.5%
Carbon monoxide	CO	Trace	0.9%
Oxygen	O_2	15%	18%
Hydrogen	H_2	Nil	Nil
Methane	CH_4	Nil	Nil
Sulfur dioxide	SO_2	80/90 ppm	192 ppm
Chlorine	Cl_2	Trace	2.5 ppm
Hydrocyanic acid	HCN	Trace	3.0 ppm
Alcohol vapors	CH_3OH, C_2H_5OH	Nil	Nil
Phosgene	$CO\,Cl_2$	Nil	Nil
Arsine	AsH_3	Nil	Nil
Mercury	Hg	Nil	Nil
Ammonia	NH_3	Nil	Nil

there is almost always a visible vapor plume at the top of the stack.[7] High energy requirements and corrosion are problems associated with wet scrubbers in this application.

20.6 USE OF ELECTROSTATIC PRECIPITATORS ON MUNICIPAL INCINERATORS

Electrostatic precipitators are used extensively throughout Europe for control of emissions from municipal incinerators. Several installations have been made in the United States and Canada within recent years, as shown in Table 20.6.

Because of the high temperature of the gases from the incinerator furnace, some means of gas cooling is required. In some installations waste heat boilers are incorporated either as an integral part of the furnace or as a separate unit. Where heat recovery equipment is not used, cooling towers equipped with water sprays are utilized for control of the inlet gas temperature.

The design of the spray cooling towers is critical to the successful performance of the precipitators. The primary requirements are that there be complete evaporation of the water in the chamber with no moisture carryover to the precipitator. Otherwise, the dust will tend to cake or form a slurry, which will give rise to buildup within the cooling tower and also present dust removal problems.

There are a number of alternative designs for cooling towers used by various manufacturers. The principal variations are the location of the gas inlet, location of the water sprays, pressure of the water to the sprays, and the method of controlling the quantity of water to the cooling tower.

Factors influencing the evaporation of water in the tower are the temperature of the inlet gas, size of the water droplets, and the time of contact of the water with the gases. Temperature of the gas from an incinerator furnace varies with the type and quantity of refuse being burned. Since the heat transfer rate is directly proportional to gas temperature, it follows that lower gas temperatures require longer contact times or greater surface area for complete evaporation.

The heat transfer rate is also related to the total surface area exposed to the gas. Thus, the smaller the droplet size, the greater the

Table 20.6

Electrostatic Precipitators Installed on Municipal Incinerators in North America

Location	Collector Manufacturer
Refractory Units:	
City of Stamford, Conn.	Univ. Oil. Prod. (Air Correction Div.)
NYC South Shore Brooklyn	Research-Cottrell
NYC Southwest Brooklyn	Wheelabrator/ Lurgi
Dade County, Florida	Wheelabrator/ Lurgi
Water Wall:	
City of Montreal, Quebec (Von Roll)	Research-Cottrell
City of Braintree, Mass. (Detroit Stoker)	Wheelabrator/ Lurgi
City of Hamilton, Ontario (C &E Boiler)	Wheelabrator/ Lurgi
City of Chicago, Illinois	Wade/ Rothemuhle
Eastman Kodak Rochester, N. Y. (suspension burning)	Wheelabrator/ Lurgi

surface area for a given water volume. Water sprays used in cooling towers are of the hydraulic atomization type, since gas velocities in the chambers are low. In general, the size of the water droplet from a hydraulic atomizer is inversely related to the hydraulic pressure. However, the size reduction decreases rapidly after a given pressure is reached. Water pressures vary with each manufacturer and range from 30 to 600 psi.

The time that the water droplet is in contact with the gas stream is related to the size of the cooling tower and the volume of gas handled. Whitehead and Darby[9] made a study of the flow patterns in cooling towers and point out that velocity gradients within the cooling towers can reduce the effective contact time. Consequently, gas flow characteristics within the tower must be taken into consideration in cooling tower design.

Control of water to the cooling tower can be achieved by grouping sprays in banks under thermostatic control to bring in the number of sprays required to maintain the gas temperature in the desired range. This type of step control is used on many types of cooling towers.

An alternative system is the use of the "spill back" principal, in which water is supplied to the spray nozzles through concentric pipes. The nozzles are of the type to provide adequate atomization while varying the throughput, thus giving a continuously variable temperature control. Continuous flow through the water supply system also provides for cooling of the piping.

The interior of the spray cooling chamber is equipped with refractory lining, as is the ductwork leading from the incinerator furnace. The linings can be made of refractory brick or of a gunned refractory mix.

Since the precipitator and fan can be damaged by exposure to the temperature of the exit furnace gases, provision should be made to bypass the gases around the precipitator in case of failure of the water supply or other components of the system.

Figure 20.3 is a schematic diagram of a complete system showing the cooling tower, precipitator, and emergency gas bypass system.

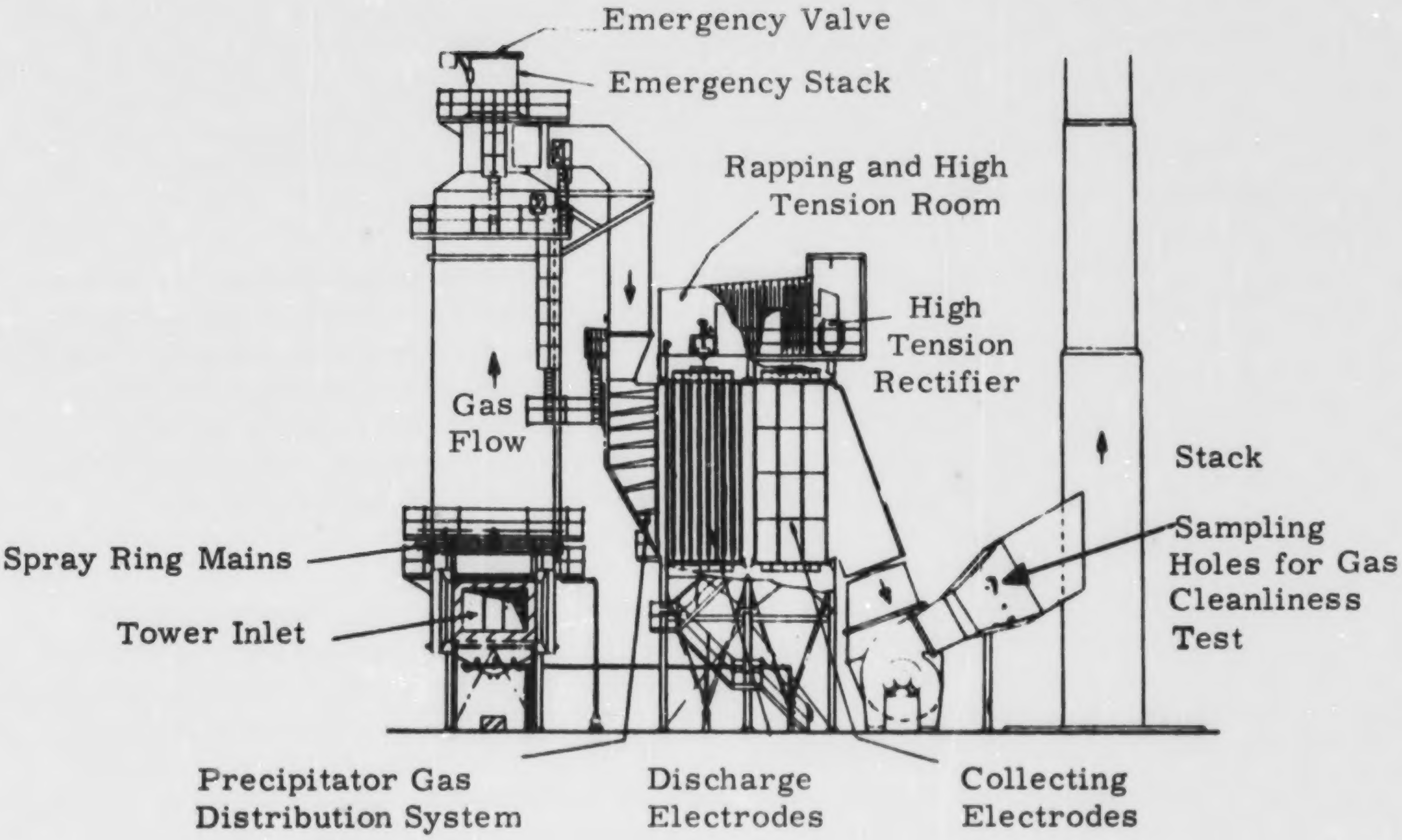
Emergency Valve
Emergency Stack
Rapping and High Tension Room
High Tension Rectifier
Gas Flow
Stack
Spray Ring Mains
Sampling Holes for Gas Cleanliness Test
Tower Inlet
Precipitator Gas Distribution System
Discharge Electrodes
Collecting Electrodes

Figure 20.3 General Layout of Electrostatic Gas Cleaning Plants for Municipal Incinerators (Reference 9)

Electrostatic precipitators used in the collection of fly ash from waste incinerators are generally of the single-stage, duct type with horizontal gas flow. Insulation is generally required on the shell to minimize corrosion due to condensation of the incinerator gases. The dust removal systems are from pyramidal or flat draft type hoppers. Discharge electrodes are of the weighted wire types in the case of the American precipitators or the supported frame electrodes in the case of the European designs. Collection electrodes and electrification equipment can be of several types, depending upon the manufacturer. These are described in Chapter 10, Part I. Rappers can be of the impact type, either electromagnetic, mechanical, or pneumatic. In addition to conventional rappers, water sprays are sometimes installed under the precipitator roof to wash down the electrodes.

Size and power requirements for precipitators utilized in collection of dust from municipal incinerators are determined from past experience on similar service. The collection surface area requirements are related to the efficiency, gas volume, and precipitation rate parameter by the Deutsch-Anderson equation. Plots of the specific collection area and power rate as a function of collection efficiency for a few installations are shown in Figures 20.4 and 20.5. A range for the precipitation rate parameter and power density is shown in Figure 20.6.

The value of precipitation rate parameter varies according to the size distribution of the particles being collected, the resistivity of the collected dust, the density of the individual particles, and the particle composition. These in turn depend on the nature of the refuse being burned. Carbon, resulting from the combustion of paper, is particularly difficult to precipitate since it has low electrical resistivity and can lose its charge on contact with the collection electrode and be reentrained. This factor, coupled with variations in dust particle size and gas composition, results in values of precipitation rate parameter over a reasonably wide range.

Values of precipitation rate parameter for municipal incinerator precipitators in European installations vary from around 4-10 cm/sec. Data on U. S. installations are insufficient to present statistically meaningful values of w.

Table 20.7 below is a summary of available data on a single precipitator installation on a municipal incinerator showing some important parameters.

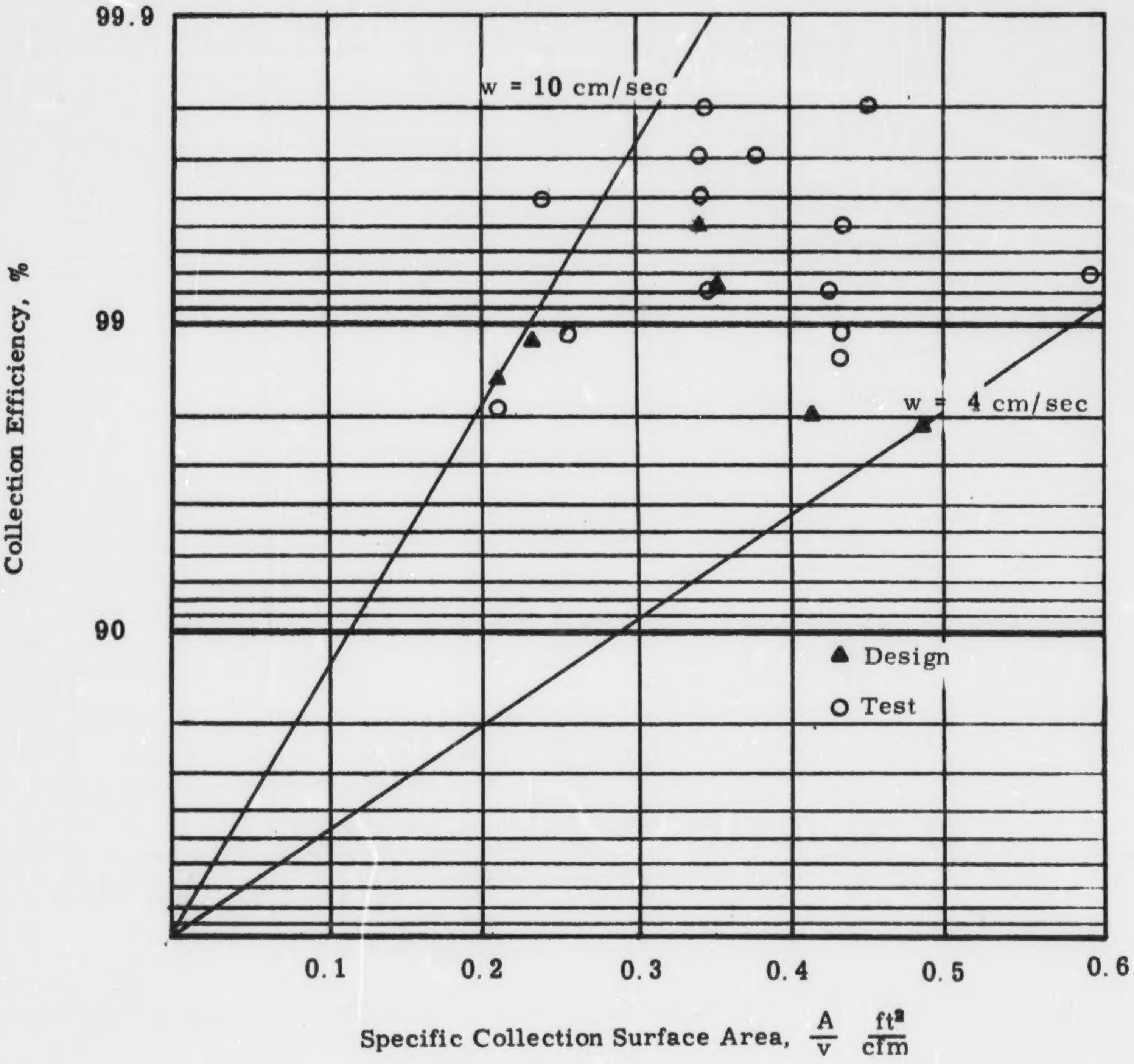

Figure 20.4. Relationship Between Collection Efficiency and Specific Collection Area for Municipal Incinerators.

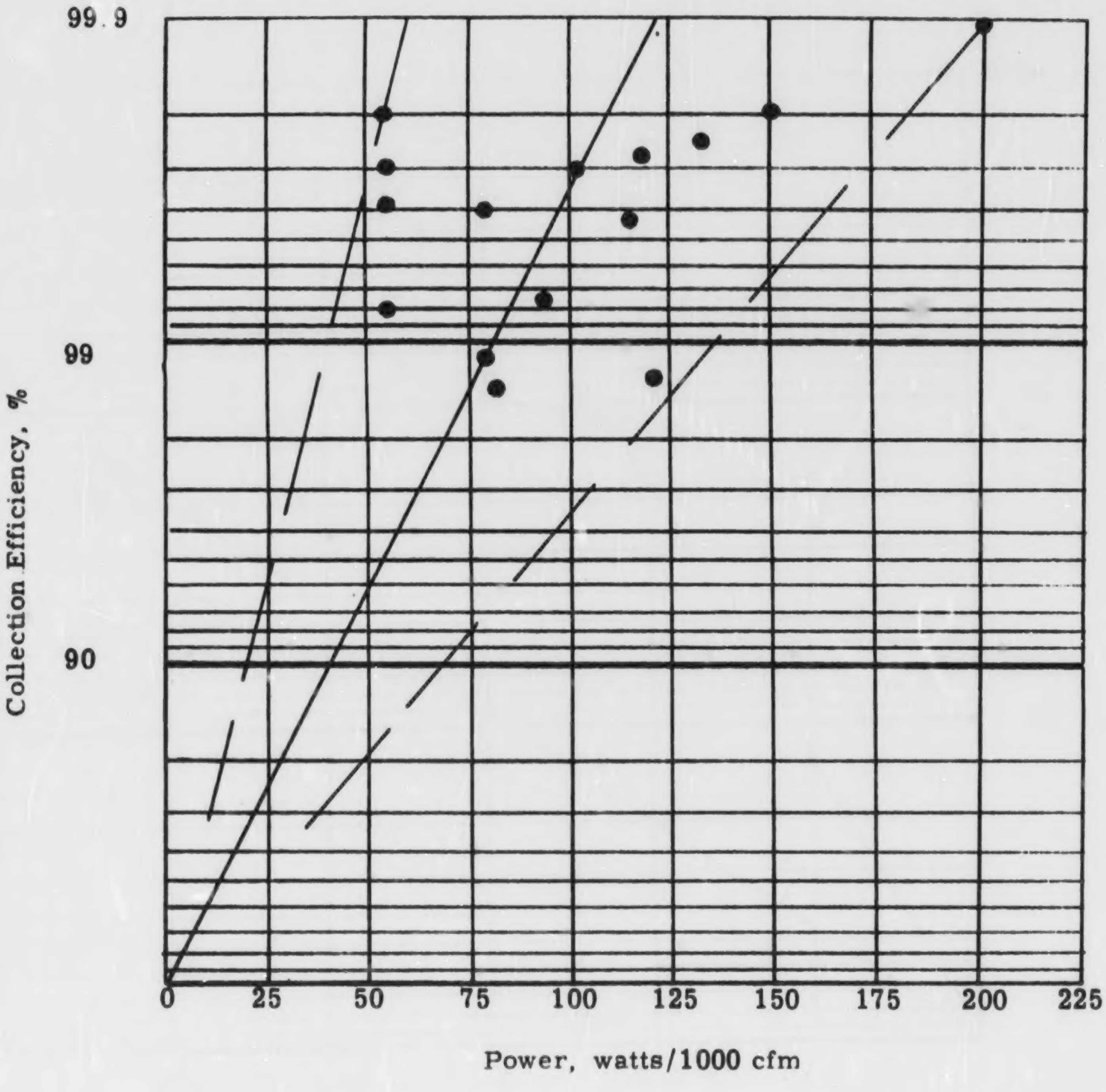

Figure 20.5. Relationship Between Collection Efficiency and Delivered Corona Power for Municipal Incinerators.

Note: Delivered power = secondary voltage x secondary current.

Input power $= \dfrac{\text{delivered power}}{\text{power supply efficiency}}$

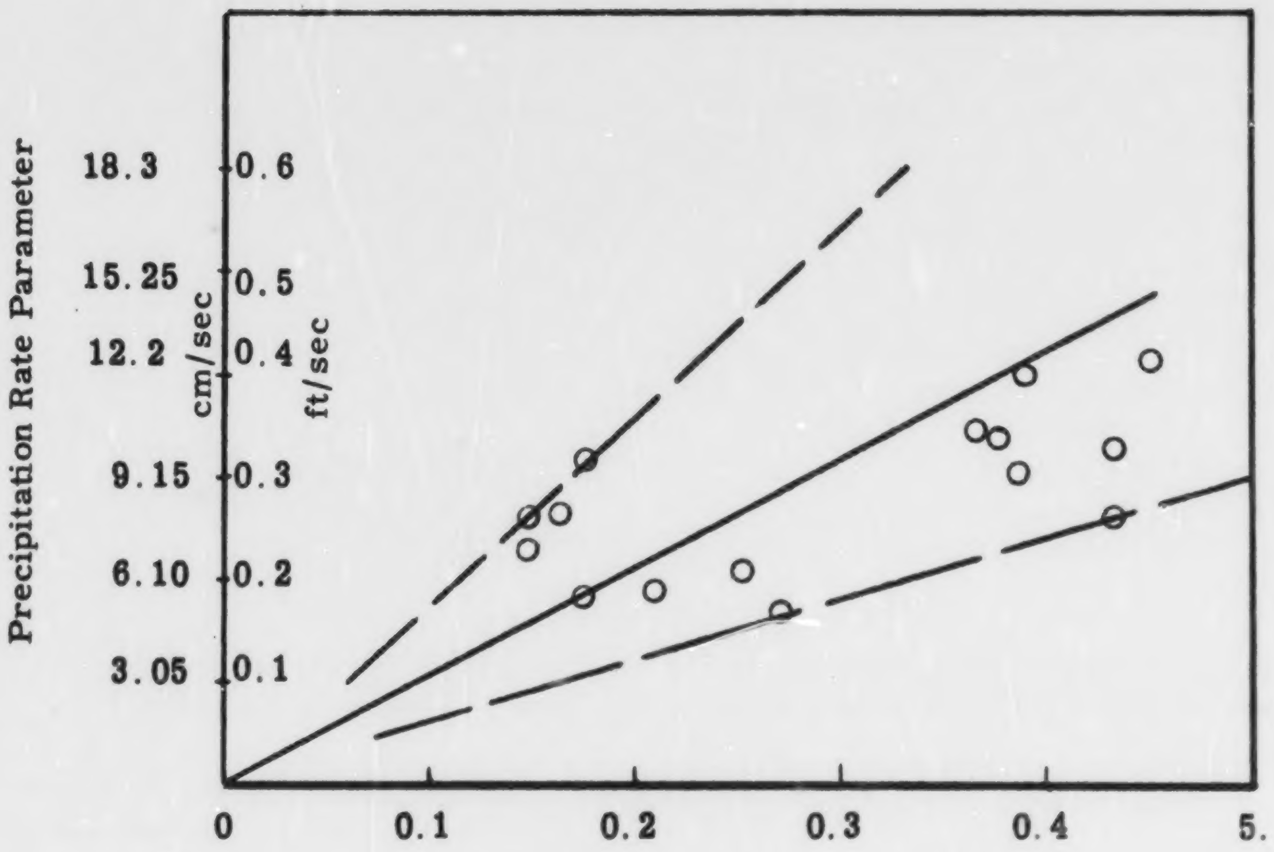

Figure 20.6. Relationship Between the Power Density and Precipitation Rate Parameter for Electrostatic Precipitators Operating on Effluents from Municipal Incinerators.

Table 20.7

Municipal Incinerator Installation

Conditioning Tower	
Water circulation rate	135 gpm
Water evaporation loss	90 gpm
Water pressure	600 lb/in.^{2}
Tower dimensions	16 ft, 9 in. dia x 42 ft, 12 in. high
Precipitator	
Gas volume	135,000 acfm
Collecting electrode area	19,500 ft^2
Gas velocity	4.6 ft/sec
Cross-sectional area	486 ft^2
No. gas passages	27
No. fields	1
Length of field	14 ft
No. of plate rappers	28
No. of discharge electrode rappers	27
Power supply	1-700 mA - 1-500 mA, 45 kV
Efficiency	95%
Gas temperature	550 - 600°F

The ranges of erected costs for 93-95% collection efficiency units, with capacities of 64,000-569,000 acfm, were 0.44 to 0.66 \$/acfm (1968).

No data are available for United States installations. Table 20.8 is a summary of performance data obtained on German incinerator installations operating on refuse, refuse and coal, or refuse and oil.

Figure 20.7 indicates precipitation rate parameters as a function of gas temperature. The variation in precipitation is probably due to the change in resistivity of the dust more than any other single factor. The spread in the data can be due to the variation in resistivity due to dust and gas composition as well as gas velocity variations and variations in precipitator design and condition of repair.

Table 20.8

Summary of Performance Data

Plant/Unit	Plant 1		Plant 2			Plant 3						
Test Number	1	2	1	2	3	1	2	3	4	5	6	7
Firing Mode	Coal Only		Coal and Refuse		Refuse Only	Coal	Coal	Coal (Low Load)	Coal-40 TPH Refuse	Coal-40 TPH Refuse	Coal-40 TPH Refuse	Coal-40 TPH Refuse
Rated Gas Volume, ft^3/sec Gas Temp °F	163.5 284°	163.5 284°	2200 320°	2200 320°	- -	5720 302°	5720 302°	5720 302°	7200 338°	7200 338°	7200 336°	7200 338°
Actual Gas Volume, 1000 x acfm Gas Temp °F	104 247°	108 257°	140 310°	140 310°	103 315°	345 308°	348 313°	244 284°	418 324°	432 336°	384 326°	403 328°
Percent of Rating, %	112	115	111	111	-	101	101	71.3	96.8	100	88.7	93.5
Anticipated Pptr. Inlet Dust Conc., gr/scf	1.97	1.97	1.97/6.95	1.97/6.95	6.95	2.19-8.75	2.19-8.75	2.19-8.75	2.19-8.75	2.19-8.75	2.19-8.75	2.19-8.75
Actual (Test) Pptr. Inlet Dust Conc., gr/scf	2.39	2.44	5.12	8.64	6.76	1.18	1.51	9.51	3.06	3.06	3.24	1.66
Actual (Test) Pptr. Outlet Dust Conc., gr/scf	0.0105	0.0325	0.0128	0.0178	0.00774	0.0069	0.0166	0.0060	0.0108	0.00503	0.00696	0.0133
Guaranteed Collection Efficiency (Corrected for Actual Test Conditions per Manufacturer's Correction Factors), %	97.94	97.49	99.25	-	-	97.97	98.00	99.55	99.55	99.50	99.72	99.54
Actual (Test Collection Efficiency), %	99.56	98.67	99.75	99.79	99.89	99.24	98.90	99.37	99.65	99.84	99.72	99.20
Pptr. Design Gas Velocity at Rated Volume (v), ft/sec	2.48	2.48	3.35	3.35	-	3.15	3.15	3.15	3.98	3.98	3.98	3.98
Pptr. Actual (Test) Gas Velocity (v), ft/sec	2.77	2.86	3.714	3.714	2.73	3.19	3.21	2.26	3.86	3.99	3.53	3.72
Relative pptr. Size, Design Based on rated flow, sec/ft $\frac{A}{V}$	29.80	29.80	22.07	22.07	-	25.41	25.41	25.41	20.18	20.18	20.18	20.18
Relative pptr. Size, Actual Based on actual flow, sec/ft $\frac{A}{V}$	26.68	25.88	19.91	19.1	27.10	25.22	25.05	35.64	20.83	20.16	22.74	20.58
Design pptr. Performance w (w design), ft/sec	0.130	0.124	0.222	-	-	.153	.154	.213	.268	.263	.291	.267
Actual pptr. Performance w (w actual), ft/sec	0.203	0.167	0.301	0.323	0.251	.193	.180	.142	.271	.319	.256	.235
Pptr. Electrical Energization Data												
A) Secondary Kilovolts Inlet (kV) (Inlet/Outlet)	41.2/44.4	40.8/43.7	30/34	31/34	32/33	38.2/37.2	37.2/37.8	-	32.2/32.6	30.5/32.2	32.2/32.2	32.2/32.8
B) Secondary Milliamps (mA) (Inlet/Outlet)	260/306	240/381	600/560	640/650	640/650	850/940	585/775	-	750/720	735/870	600/660	620/684
C) Input Power (AxB) (Kilowatts) (Inlet/Outlet)	10.7/13.7	9.79/16.6	18.0/19.0	19.8/22.1	20.5/21.4	26.5/34.9	21.8/29.3	-	24.2/23.5	22.4/28.0	19.3/21.3	20.0/22.4
D) Power Density-Watts per 1000 acfm (Inlet/Outlet)	103/126	90.6/154	128.6/136	141.7/157.9	198.8/206	76.9/101	62.5/84.2	-	57.8/56.2	51.9/64.8	50.3/55.3	49.5/55.7
E) Power Density-Watts per ft^2 (Inlet/Outlet)	.220/.281	.201/.342	.370/.391	.407/.454	.421/.440	.183/.241	.150/.202	-	.167/.162	.155/.193	.133/.147	.138/.155
F) Field Strength-Kilovolts per inch (Inlet/Outlet)	0.87/0.94	0.86/0.92	.63/0.72	0.65/0.72	0.68/0.70	0.80/0.77	0.77/0.80	-	0.68/0.69	0.65/0.68	0.68/0.68	0.68/0.69
G) Sparking Intensity/Frequency	-	-	-	-	-	-	-	-	-	-	-	-

Notes: 1) All tests, unless noted, performed at full boiler steaming load.

2) All efficiency data calculated on basis of dust loadings at (0°C - 760 mm Hg)

3) Precipitator sizing calculated from original metric units; recalculations in English units will compound rounding-off errors from unit-conversions.

Table 20.8 (Continued) - Summary of Performance Data

Plant/Unit	Plant 4		Plant 5			
Test Number	1	2				
Firing Mode	Refuse	Refuse	Refuse and Oil	Refuse and Oil	Refuse and Oil	Refuse and Oil
Rated Gas Volume, ft^3/sec	1550	1550	2870	2870	2870	2870
Gas Temp °F	500°	500°	410°	410°	410°	410°
Actual Gas Volume, 1000 x acfm	91	92	156	127	161	111
Gas Temp °F	455°	468°	375°	375°	362°	360°
Percent of Rating, %	97.7	98.9	95.2	77.4	98	67.6
Anticipated Pptr. Inlet Dust Conc., gr/scf	3.94	3.94	1.81	1.81	1.81	1.81
Actual (Test) Pptr. Inlet Dust Conc., gr/scf	4.81	5.69	1.67	1.83	1.47	1.10
Actual (Test) Pptr. Outlet Dust Conc., gr/scf	0.0158	0.0184	0.0169	0.0207	0.0210	0.00283
Guaranteed Collection Efficiency (Corrected for Actual Test Conditions per Manufacturer's Correction Factors), %	98.85	98.95	98.5	99.5	97.9	99.5
Actual (Test Collection Efficiency), %	99.67	99.68	98.22	98.87	98.57	99.74
Pptr. Design Gas Velocity at Rated Volume (v), ft/sec	3.82	3.82	3.68	3.68	3.68	3.68
Pptr. Actual (Test) Gas Velocity (v), ft/sec	3.74	3.77	3.51	2.85	3.61	2.49
Relative pptr. Size, Design $\frac{A}{V}$ Based on rated flow, sec/ft	14.00	14.00	12.17	12.17	12.17	12.17
Relative pptr. Size, Actual $\frac{A}{V}$ Based on actual flow, sec/ft	14.32	14.16	12.79	15.73	12.42	18.02
Design pptr. Performance w (w design), ft/sec	.408	.319	.328	.435	.317	.435
Actual pptr. Performance w (w actual), ft/sec	.399	.406	.315	.285	.342	.330
Pptr. Electrical Energization Data						
A) Secondary Kilovolts Inlet (kV) (Inlet/Outlet)	31.5/29	31/29	-	-	26.5/21.8	26.8/22.4
B) Secondary Milliamps (mA) (Inlet/Outlet)	265/267	313/310	-	-	484/588	503/575
C) Input Power (AxB) (Kilowatts) (Inlet/Outlet)	8.3/7.7	9.7/9.0	-	-	12.8/12.8	13.5/12.9
D) Power Density-Watts per 1000 acfm (Inlet/Outlet)	91.7/85	105/97.7	-	-	79.7/79.6	121/116
E) Power Density-Watts per ft^2 (Inlet/Outlet)	.401/.372	.466/.432	-	-	.366/.366	.385/.368
F) Field Strength-Kilovolts per inch (Inlet/Outlet)	0.74/0.68	0.73/0.68	-	-	0.61/0.50	0.61/0.51
G) Sparking Intensity/Frequency	-	-	-	-	-	-

Notes: 1) All tests, unless noted, performed at full boiler steaming load.

2) All efficiency data calculated on basis of dust loadings at (0°C - 760 mm Hg)

3) Precipitator sizing calculated from original metric units . . . recalculations in English units will compound rounding-off errors from unit-conversions.

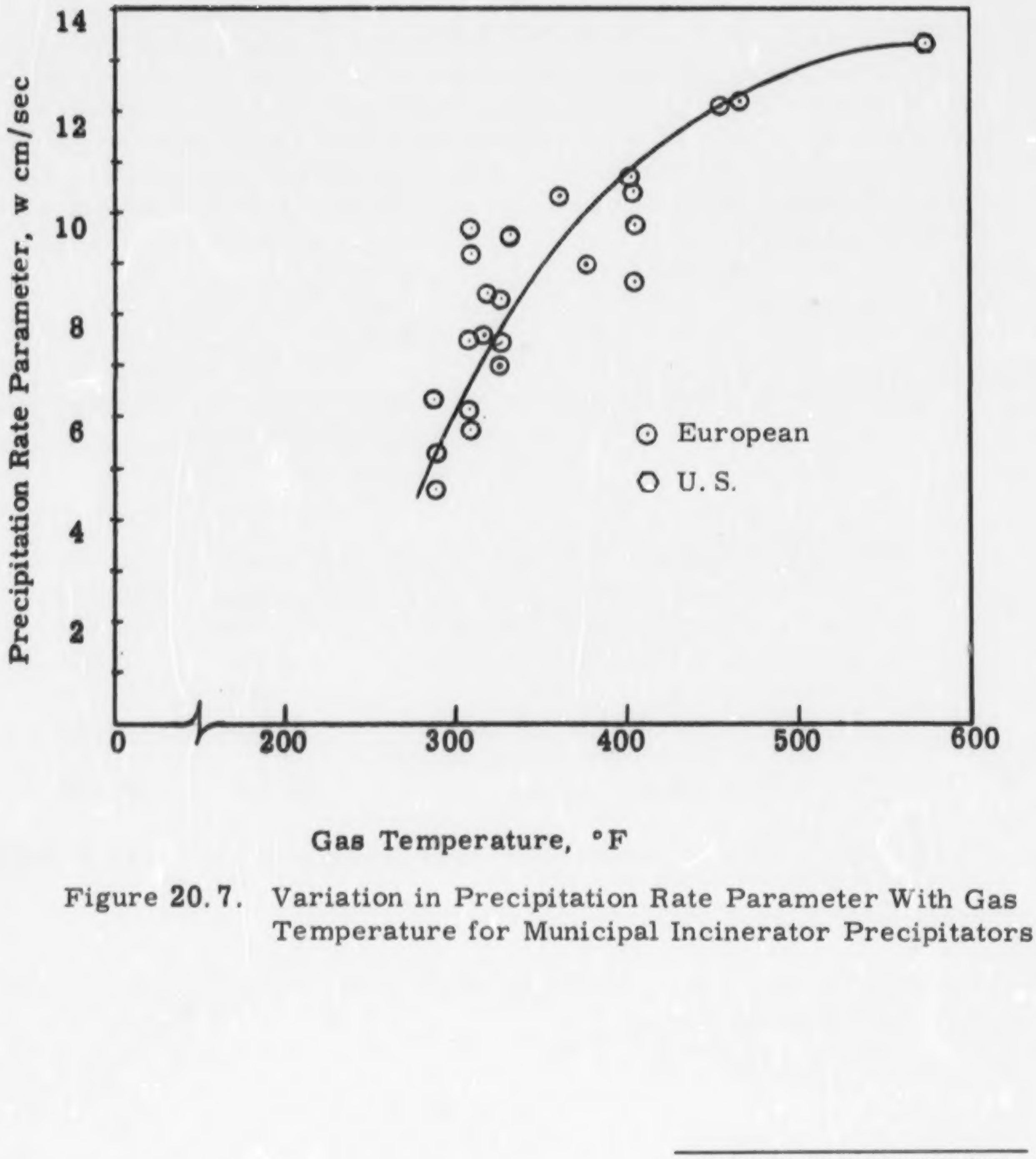

Figure 20.7. Variation in Precipitation Rate Parameter With Gas Temperature for Municipal Incinerator Precipitators.

20.7 ECONOMICS

Cost data on U. S. installations are limited because of the limited number of installations. The two installations for which costs are available range from $1.68/acfm to $2.90/acfm. These are "turnkey" costs (1968) for 95% efficiency. Costs include pilings, foundations, supports, flues, heat insulation, piping, and wiring as well as the precipitator and power supply. The costs do not include buildings.

Operating cost data for the two installations are not available since they have not been in operation for sufficient time to accumulate data. However, operating costs can be estimated based on power requirements and capital costs assuming maintenance costs consistent with other precipitator applications. Table 20.9 shows estimated operating costs for two size precipitators. For this estimate, capital charges are based on 12½% of the installed costs. Maintenance costs are computed on the basis of 1% of the installed costs per year.

Table 20.9

Estimated Operating Costs for Electrostatic Precipitators for Municipal Incinerator Dust Collection

	1	2
Gas Flow (acfm)	250,000	130,000
Installed Cost ($ 1968)	420,000	390,000
Capital Charges (12.5%)	52,500	48,800
Maintenance Costs (1% of cap)	4,200	3,900
Precipitator Energy[1] Costs	605	336
Fan Energy Costs[2]	1,760	923
Total Yearly Operating Costs 1 shift (2000 hr) basis	59,105	53,959

[1] Computed on basis of precipitator input power (see Figure 20.5) 2000 hrs opr/year (1 shift) and 1.5 cents/kwh energy costs and 60% power supply efficiency.

[2] Computed on basis of 1" water pressure drop, 50% efficiency for the fan and motor combination.

$$\text{Fan energy} = \frac{v \times p \times 746}{6356 \times \text{eff}} \times 10^{-3}$$

where

v = gas volume acfm,
p = pressure in. water,
eff = fan and motor efficiency, and
fan energy = kW.

CHAPTER 20
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CHAPTER 21
ELECTROSTATIC PRECIPITATORS IN THE PETROLEUM INDUSTRY

21.1 INTRODUCTION

The principal uses of electrostatic precipitators in the petroleum industry are for the collection of particulate emission from fluidized bed catalytic cracking units and removal of tar from various gas streams, such as fuel gases, acetylene, and shale oil distillation gases.

The first of these areas, recovery of catalyst dust, originated with the production of high octane gasoline in fluid catalytic cracking units during World War II. In order to economically operate these units, electrostatic precipitators were used to recover catalyst from the discharge stream of the catalyst regenerators. While improvement of mechanical collectors inside the regenerators has eliminated the process requirement for electrostatic precipitators, precipitators are presently used as control devices for the recovery of catalyst fines produced by attrition of the catalyst.

The second major application of electrostatic precipitation is encountered in the detarring of gases produced in a variety of processes. The gasification of organic material for the production of these gases results in the formation of small amounts of tars and oils which must be removed prior to further use. Similar applications exist for the related area of solid fuel gasification.

The precipitation of tars from fuel gases is technically straightforward inasmuch as the precipitated tars are free flowing, and steady operating conditions can be maintained without the rapping and reentrainment problems associated with dry precipitators. Since the sparking inherent in precipitator operation would ignite the gas, the gas-oxygen ratio must be maintained outside combustion limits to eliminate the fire or explosion hazard. Fuel-air ratio sensors operate automatic control equipment to provide this safeguard. Precipitators for detarring applications are reasonably trouble free, except for occasional corrosion problems. Installation costs and operating efficiencies are so favorable for this application that other techniques for detarring are not generally considered.

A relatively new application of electrostatic precipitators is in the removal of tar, fine carbon, and oil mist from the acetylene gas manufactured from naptha or crude oil.

Another relatively new application in the petrochemical field is the utilization of precipitators for the removal of tars from oil shale distillation gases. Electrostatic precipitators have been successfully employed in this application for cleaning the gases which are subsequently condensed and treated to produce petrochemical products.

Each of these application areas will be discussed in more detail in subsequent sections.

21.2 CATALYTIC CRACKING

In the fluidized bed catalytic cracking process, the precipitator is employed as a secondary collection system for mechanical collectors located in the top of the fluid catalyst regenerator. A typical flow diagram of a fluid cracking unit is shown in Figure **21.1.** In the production of high octane gasoline, oil and powdered catalyst are mixed in a reactor where the oil is vaporized, and the cracking reaction occurs. Spent catalyst, containing residual carbon (or coke) from the catalytic cracking process taking place in a reactor, is mixed with combustion air and fed to the regenerator to reactivate the catalyst. Reactivation consists of burning off the coke or residual carbon formed on the catalyst during the cracking process. After regeneration, the hot incinerated catalyst is mixed with crude oil and recycled to the reactor.

The gaseous products of combustion are exhausted from the top of the regenerator through a series of mechanical collectors. These mechanical collectors remove all but a very small percentage of fine catalyst and return it directly to the process. The catalyst escaping with the discharge gas may be collected in an electrostatic precipitator in order to comply with air pollution regulations. Gases leaving the regenerator may be cooled with steam or water sprays to protect down-stream equipment.

The regenerator may be followed by waste heat steam boilers that recover some of the energy and further reduce the gas temperature prior to entering the final catalyst-removal precipitator. The following parameters are typical of the process:

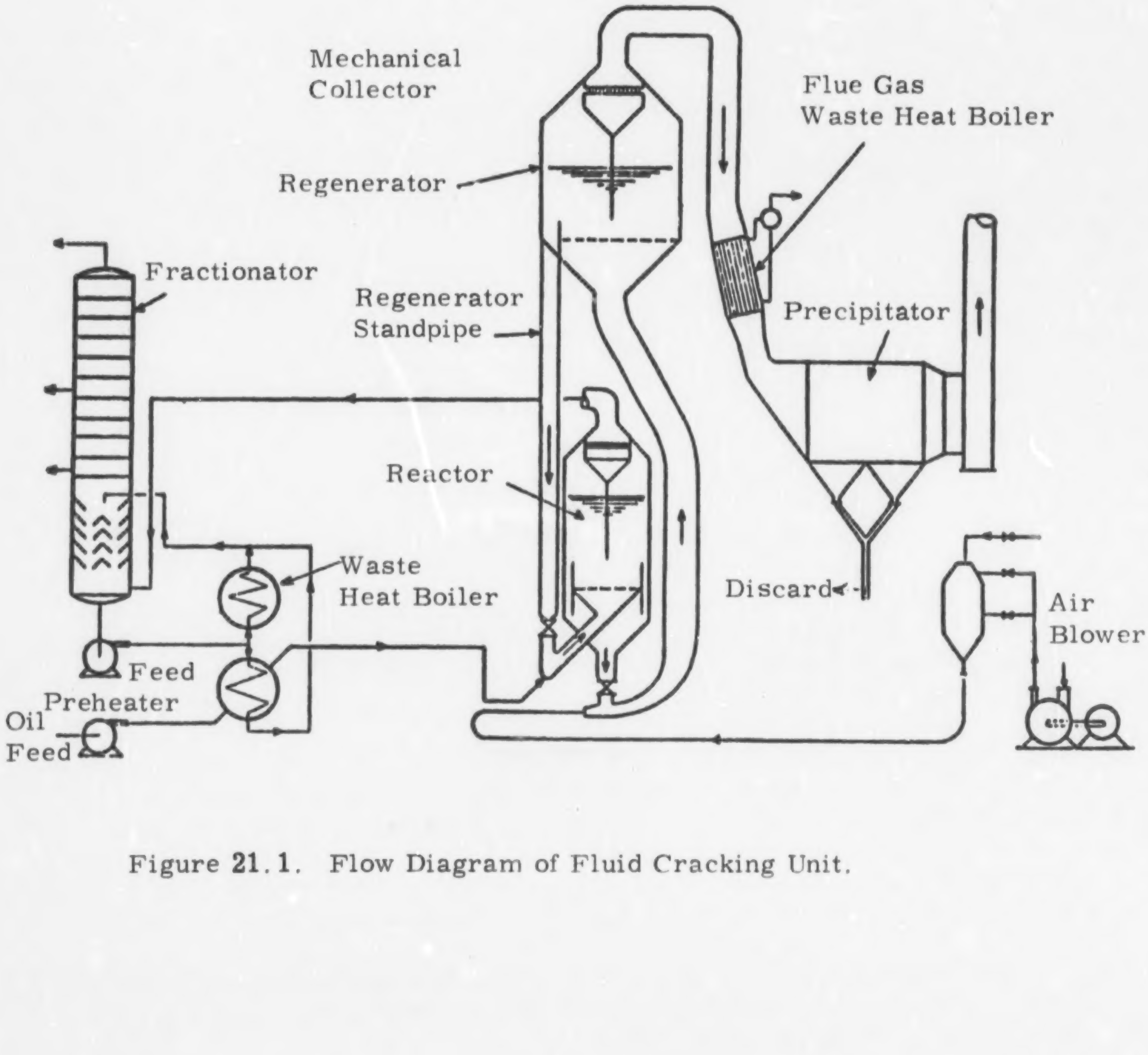
Mechanical
Collector
Flue Gas
Waste Heat Boiler
Regenerator
Fractionator
Regenerator
Standpipe
Precipitator
Reactor
Waste
Heat Boiler
Discard
Air
Blower
Feed
Preheater
Oil
Feed

Figure 21.1. Flow Diagram of Fluid Cracking Unit.

Regenerator bed temperature - 1250°F

Regenerator outlet temperature - 1100 to 1250°F

Regenerator gas pressure - 5 to 40 psig

Precipitator inlet temperature - 400° to 800°F

Dry flue gas analysis (% by volume)

N_2 = 87% - 81%

CO_2 = 6% - 9%

CO = 6% - 7%

O_2 = 1% - 3%

SO_2 = trace

Hydrocarbons = trace

Moisture content = 10% to 30% by volume

The dust concentration entering the precipitator depends on the type catalyst, the particular process, and the type of mechanical collectors used. The normal dust concentration following the mechanical collector varies between 0.02 and 1 gram per std cu ft.

The particle size is 90 - 99% less than 44μ in diameter. Figure 21.2 shows a typical particle size distribution in the precipitator inlet gas in a catalytic cracking regeneration system. The bulk density of the dust is on the order of 5 to 20 pounds per cubic foot. Oxides of alumina are the base material in most catalysts. The surface of the catalyst may be coated with other material in order to obtain maximum catalyst activity.

The typical electrical resistivity of catalyst dust measured in the laboratory for a gas temperature range of 300° to 600°F containing about 25% water vapor is shown in Figure 21.3.

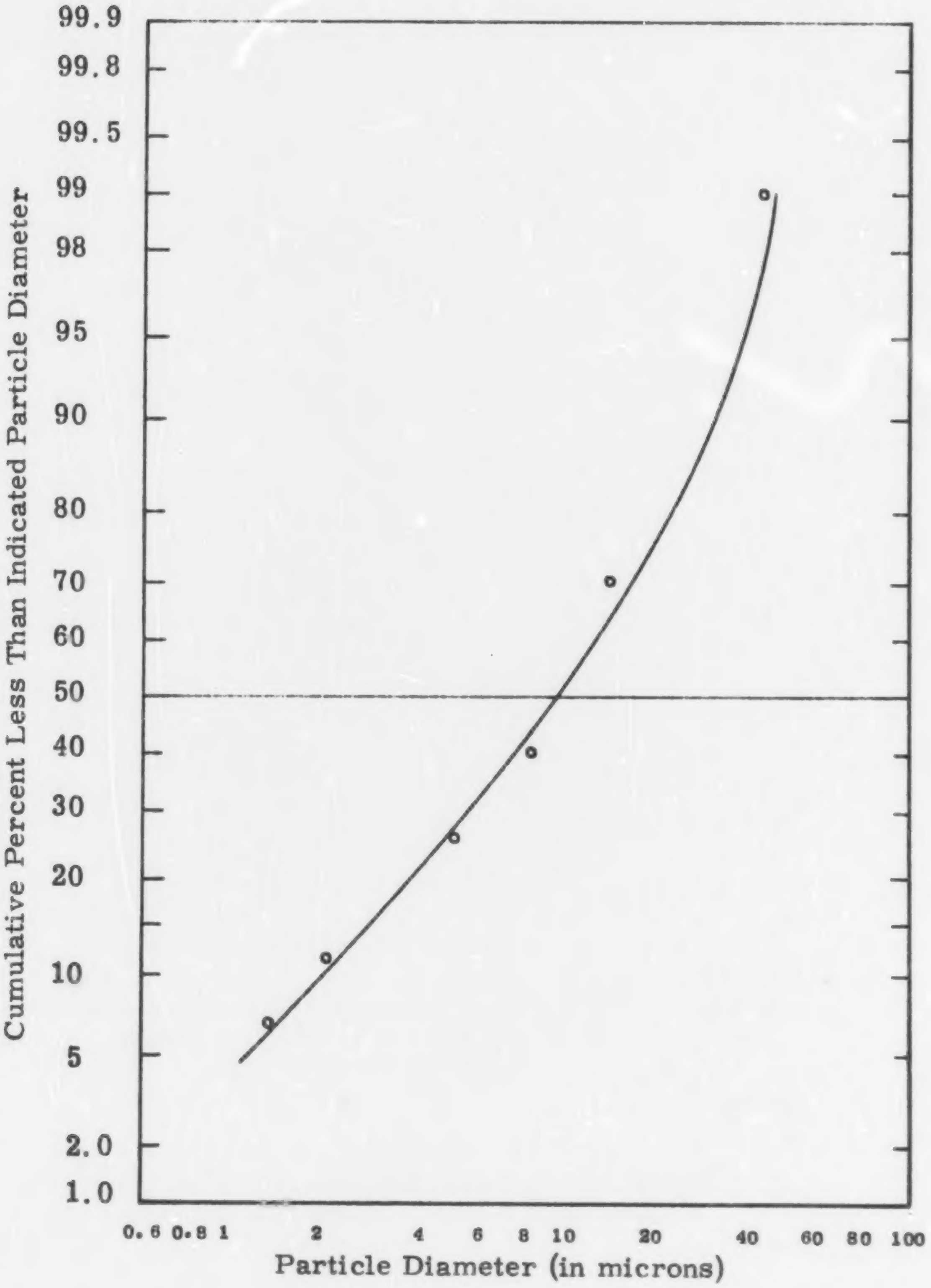

Figure 21.2. An Analysis of Particle Size Distribution in a Gas Stream to Electrostatic Precipitator for One Type of Catalyst

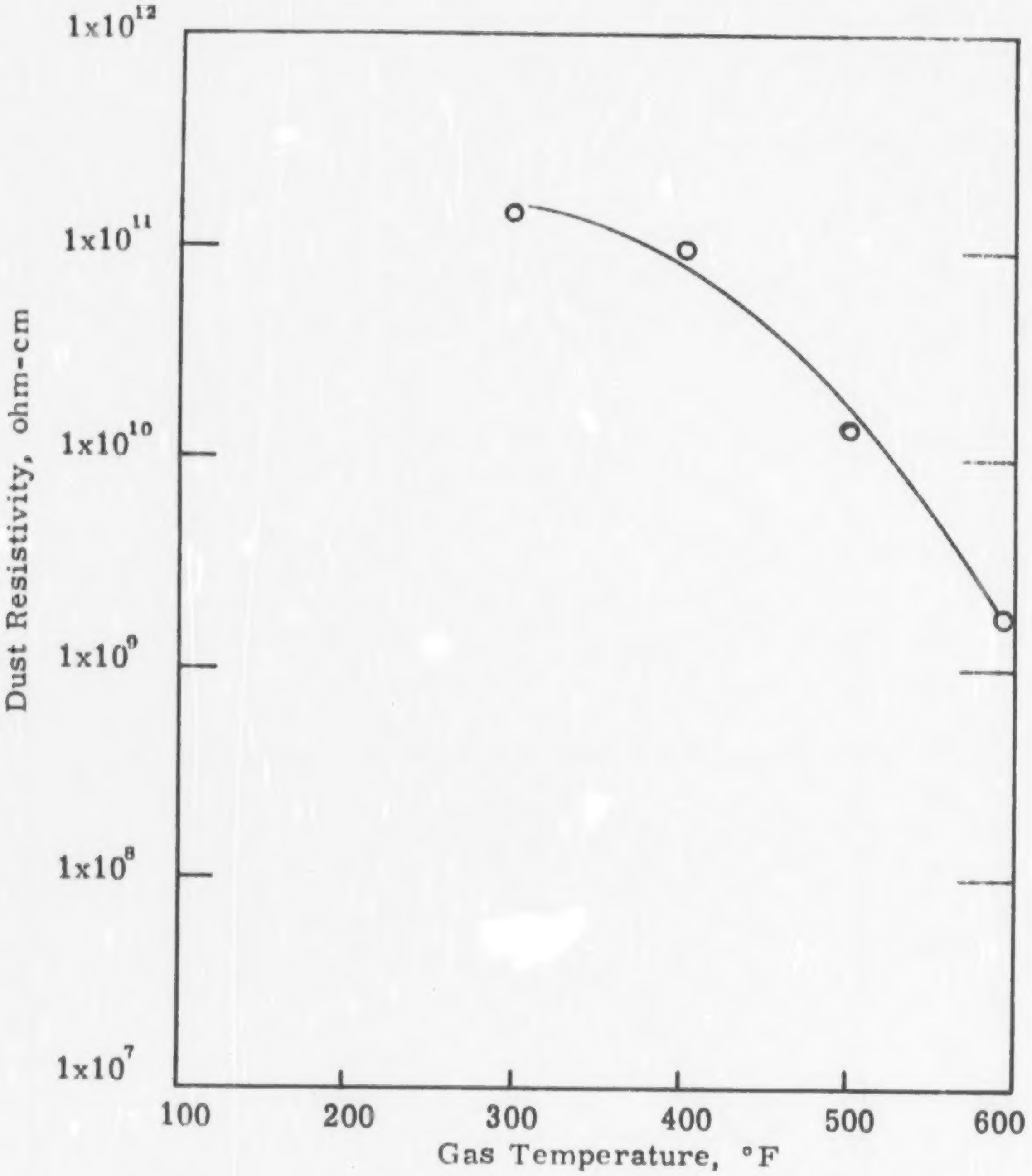

Figure 21.3. Electrical Resistivity of One Particular Type Precipitator Inlet Catalyst Dust (23% Moisture Content by Volume) - Measured in Laboratory.

In the temperature ranges in which catalyst dust precipitators operate, the electrical resistivity can be relatively high. White[1] reports that the resistivity of a typical aluminum silicate catalyst dust is around 5×10^{11} ohm-cm with no conditioning. The addition of about 20 ppm of ammonia to the gas stream reduces the resistivity to about 10^{10} ohm-cm. In the example reported by White, the precipitator efficiency increased from about 99% without conditioning to 99.8% with conditioning.

The precipitator for the collection of catalyst in the petroleum cracking process is generally a single-stage, horizontal-flow, duct type. An earlier vertical-flow design has been largely superceded by the present horizontal-flow type. The shell is constructed of steel, which is sometimes insulated on the inside with gunite or with mineral wool insulation on the outside. The fines collected are of little value and are normally discarded.

21.3 DETARRING

This discussion of detarring of gases from the petroleum industry encompasses the related areas of gasification of solid fuels as well as petroleum, since the same type of electrostatic precipitator is common to all of these operations. Minor modifications such as those found in precipitators for acetylene production are covered within the appropriate section.

The electrostatic precipitator most commonly utilized for detarring and cleaning of gaseous products is a single-stage vertical wire and pipe unit, as illustrated in Figure 21.4.

The shell is typically cylindrical, with the electrodes suspended from a top header. The high voltage support insulators for the header are mounted in external turrets which are heated to prevent moisture condensation. In some instances, these compartments are kept under continuous inert gas purge.

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[1] Refer to the bibliography for this chapter.

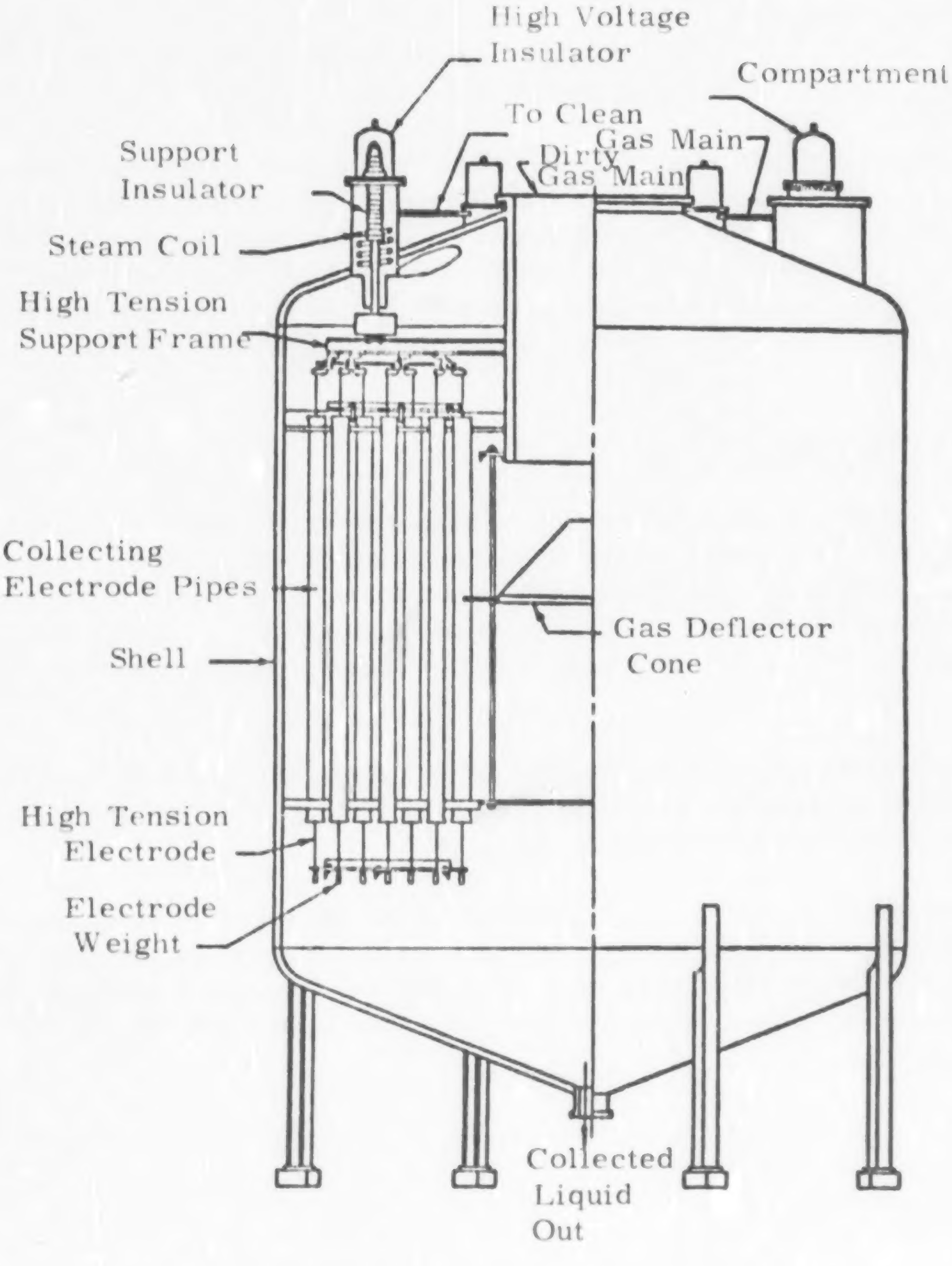
High Voltage Insulator
Compartment
To Clean Gas Main
Dirty Gas Main
Support Insulator
Steam Coil
High Tension Support Frame
Collecting Electrode Pipes
Gas Deflector Cone
Shell
High Tension Electrode
Electrode Weight
Collected Liquid Out

Figure 21.4. A Single-Stage Vertical Wire and Pipe Unit.

In most cases the collected oils and tars are free flowing, and no mechanical rapping devices are required to remove the material from the collecting electrode pipes. Removal of the accumulated drainage from a collection sump is accomplished through a liquid trap.

In some instances, the cleaning of acetylene for example, water flushing accomplishes the removal of the adherent material. This procedure is described in the section on acetylene production.

Shale oil. The process for separating oil from shale is shown in the typical flow diagram, Figure 21.5. The shale to be processed is fed into a retort where heat is applied to drive off the gas, oil, and carbonaceous residue. From the retort, the gas is passed through indirect water coolers to reduce the temperature. Most of the condensable material condenses as a submicron size mist in the cooler. The mist-laden gas is then passed through the precipitator, where the condensed oil and carbonaceous residue are separated, to the next stage for further processing. The oil and carbonaceous material collected in the precipitator sump are removed by a pump which transfers it to various locations for further separation and processing.

Acetylene. Natural gas is used as the primary raw material in the production of acetylene. The gas is burned in an oxygen-limited atmosphere under controlled temperature and pressure. The gaseous products consist of from 50 to 55% hydrogen, from 30 to 35% carbon monoxide, and from 7 to 9% acetylene. Because of the explosion hazards involved, the oxygen content is rigidly controlled to a fraction of 1%. In addition to the gases, submicron size particulate carbon is also produced and carried along by the gas as a smoke. Prior to entering the precipitator, the gas is passed through a direct-contact water cooler where the temperature is reduced to about 100° F. Some carbon is removed in the cooler. The gas then enters a series of electrostatic precipitators where practically all of the remaining carbon is removed from the gas. The precipitator is operated at a normal positive pressure of from 1 to 2 lb per square inch to prevent air in-leakage. The high voltage insulator compartments are kept under continuous purge with inert nitrogen gas. After leaving the precipitator, the cleaned gas is further processed to separate the various products. A simple process flow diagram of acetylene manufacture is shown in Figure 21.6.

Since the carbonaceous material collected is not free flowing, the inside surface of the collecting pipes is continuously flushed by a film of

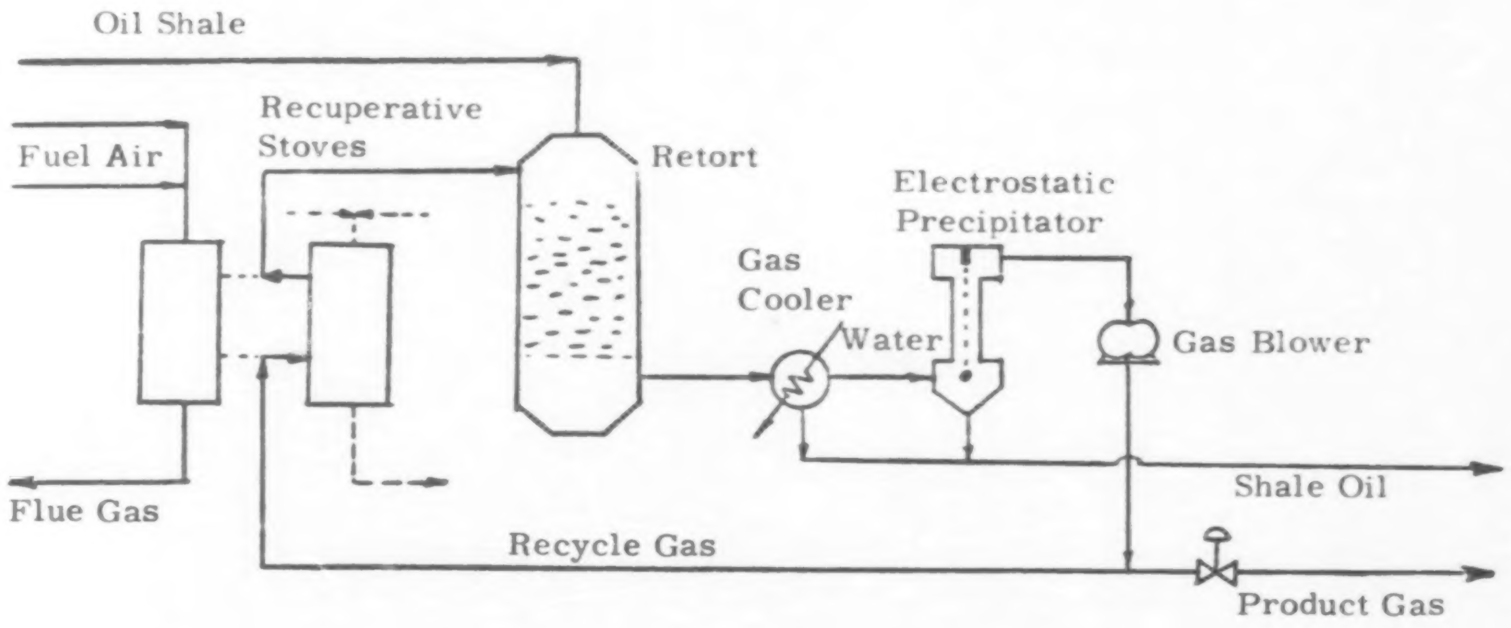
Oil Shale
Fuel Air
Recuperative Stoves
Retort
Electrostatic Precipitator
Gas Cooler
Water
Gas Blower
Flue Gas
Shale Oil
Recycle Gas
Product Gas

Figure 21.5. Oil Shale Retorting Process.

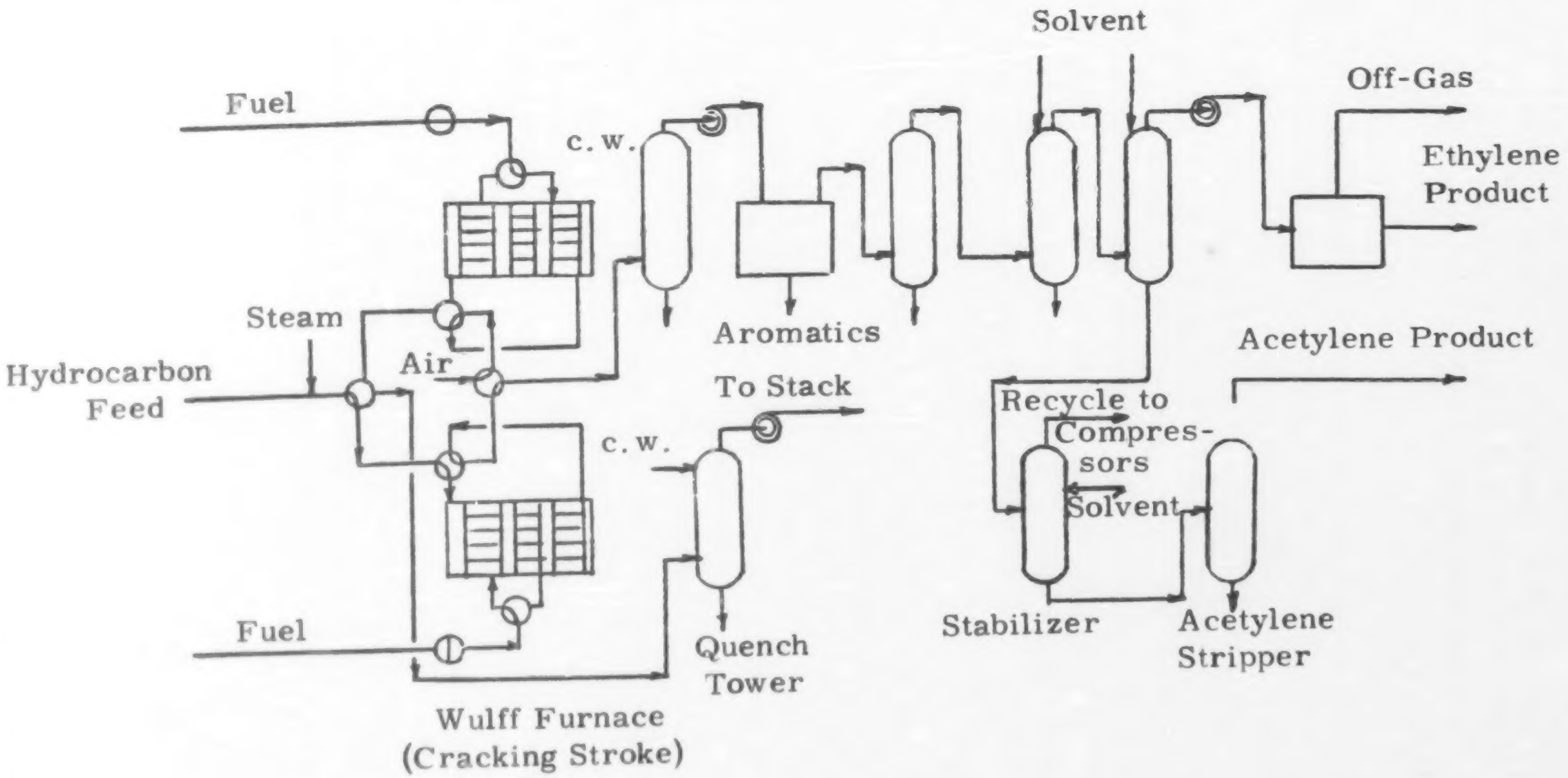
Fuel
Steam
Hydrocarbon Feed
Air
Fuel
c. w.
c. w.
Wulff Furnace (Cracking Stroke)
Aromatics
To Stack
Quench Tower
Solvent
Off-Gas
Ethylene Product
Acetylene Product
Recycle to Compres-sors
Solvent
Stabilizer
Acetylene Stripper

Figure 21.6. Acetylene (Wulff Process).

water from overflow weirs attached to the top of the pipe. A pond water supply system is provided as an integral part of the top support header. The precipitator is further provided with sprays located under the roof for additional pond and pipe cleaning. The bottom of the shell is designed as a sump to collect the water/carbon mix. Following removal, the mixture is pumped to a treatment plant where the water is separated, and the carbon pelletized.

Manufactured fuel gas. In the process for manufacturing carburetted water or producer gas, the initial gas known as "blue gas" is made by passing steam through a bed of incandescent carbon in the form of coke or anthracite coal. There are two cycles, designated in plant parlance as "make" and "blow." By means of automated regulating equipment, supervised by the gas maker and his assistant, the coal or coke is fed to the gas generator and air is admitted under the bed during the "blow" cycle. The air is then shut off and steam is admitted for the "make" cycle. The residue ash is then withdrawn. Carburetted blue gas is a mixture of blue gas (or water gas as it is often termed—made as above) and oil gas formed by the cracking of oil in a chamber through which the blue gas passes.

The carburetting process enriches the blue gas to as much as 700 Btu per cubic foot of gas, depending on the amount of oil used. With the present day use of natural gas, it is customary practice to mix natural gas with the above gases, the mixture being automatically controlled by means of calorimetric equipment. Prior to entering the precipitator, the gas is passed through a direct-contact water cooler where the temperature is reduced to around 100°F and saturated with water vapor. In the cooler, some of the tars and oils contained in the gas are removed by the water sprays, and sulfur compounds and other gaseous material are removed from the system in purifiers located after the electrostatic precipitator. A process flow diagram of production of carburetted water gas is shown in Figure 21.7.

Inasmuch as the tars and oils collected in the precipitator are free flowing, no devices are required to aid in removing the collected material from the collecting pipes. The free flowing material drains from the collecting electrode into a sump below and to an external tar pot through a liquid trap. From the tar pot, the liquid is pumped to its final point of reuse or disposal.

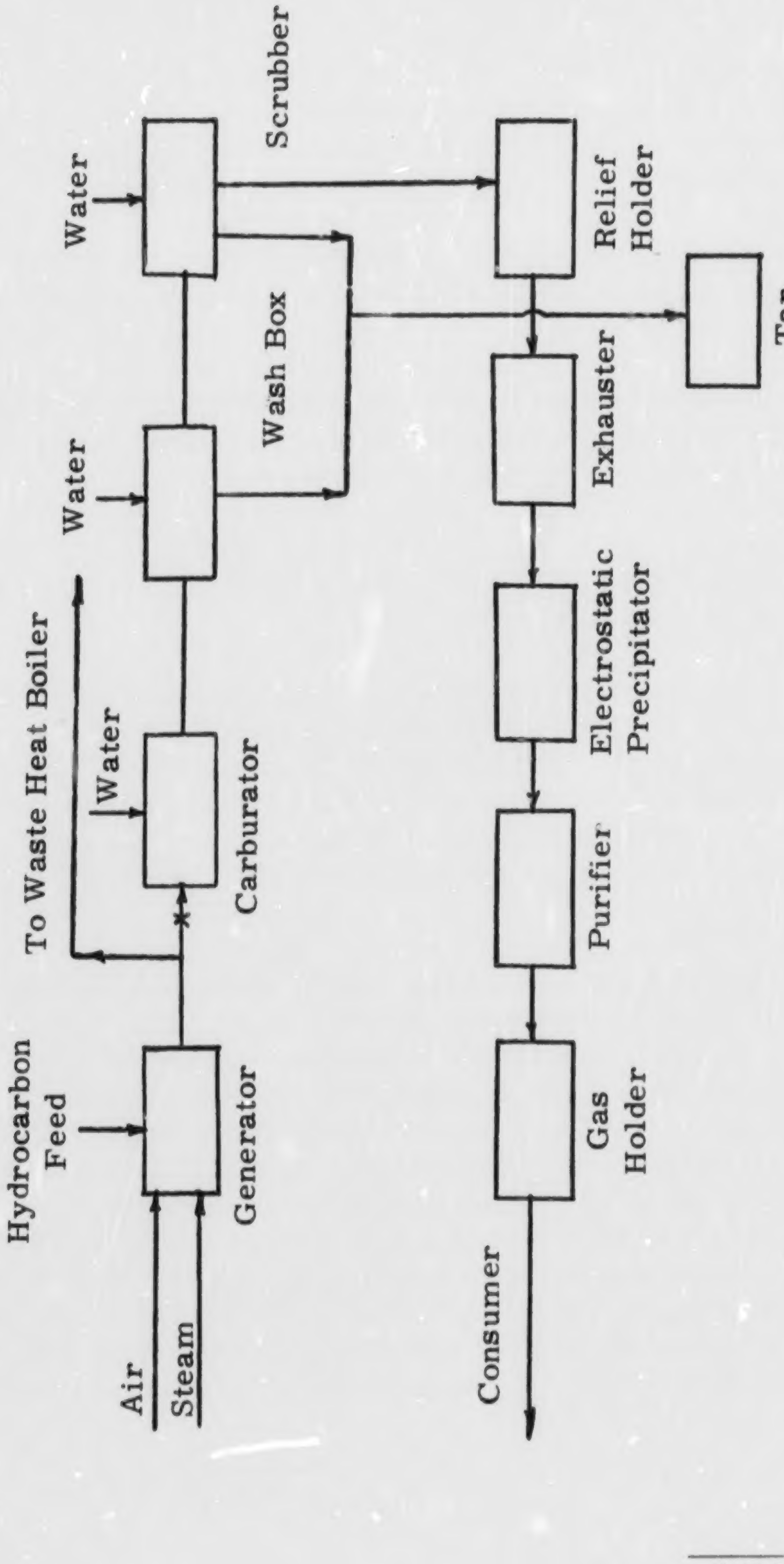
Air
Steam
Hydrocarbon Feed
Generator
To Waste Heat Boiler
Water
Carburator
Water
Wash Box
Water
Scrubber
Relief Holder
Tar Sump
Exhauster
Electrostatic Precipitator
Purifier
Gas Holder
Consumer

Figure 21. 7. A Process Flow Diagram of Production of Carburetted Water Gas.

Solid fuel carbonization. The carbonization processes, in general, consist of the heating of solid fuel materials, such as coals, lignite, and peat, to temperatures between 1020° and 1560°F in the absence of air. The fuel is subsequently decomposed into semi-coke, tar, and gas.

In addition to the coke and gas, substantial yields of liquid by-products are produced, which may be further processed to gasoline, diesel oil, fuel oil and waxes.

A typical process for carbonization is illustrated in Figure 21.8. The electrostatic precipitator is utilized for the removal of tar from the circulation and carbonizer gases leaving the carbonizing chamber.

21.4 DEVELOPMENT OF ELECTROSTATIC PRECIPITATOR IN THE PETROCHEMICAL INDUSTRY[2]

One of the earliest Cottrell electrical precipitators installed was for the cleaning of illuminating gas. It consisted of an experimental single pipe unit operated by Professor J. Davidson of the University of British Columbia in 1914 on coal and carburetted water gas at the Vancouver Gas Company. About the same time, similar experiments were carried out on producer gas at the Minnesota Steel Company at Duluth, Minnesota and at the Ann Arbor Gas Company.

In these first applications, gas was usually treated following the wash boxes or scrubbers at low temperature, so that the tar was collected in the presence of water. Shortly thereafter, in 1916, a similar precipitator was operated on carburetted water gas at about 500°F at the Tacoma Gas Company. These early experimental and semi-commercial operations soon led to regular commercial installations. In 1924, there were 5 such installations in the United States cleaning about 70 million cu ft/day, and only 8 years later there were over 75 installations handling about $1\frac{1}{4}$ billion cu ft/day. These were largely used on carburetted water gas and coke oven gas. Virtually all of the precipitators operated at temperatures from 75° to 125°F at atmospheric pressures. Typical collection efficiencies ranged from 95 to 99%. By January 1949, there were about 275 installations which included a total of 393 precipitators.

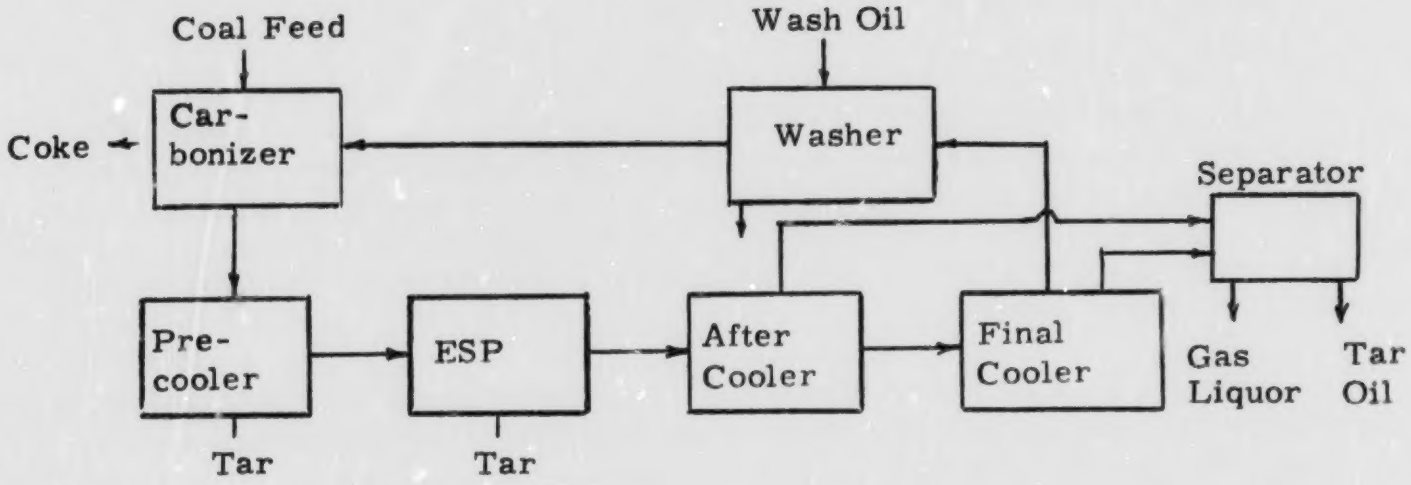
Coal Feed
Wash Oil
Coke
Car-
bonizer
Washer
Separator
Pre-
cooler
ESP
After
Cooler
Final
Cooler
Gas
Liquor
Tar
Oil
Tar
Tar

Figure 21.8. Flow Diagram for Typical Coal Carbonization Process.

By 1962, there were about 600 tar precipitators treating 5 million cfm of fuel gas. Since 1952, there have been fewer than 25 additional tar precipitators installed; some of these were replacements for the older units.

Although the average number of yearly installations has been declining, there is always the possibility of applications to entirely new fields, such as shale oil gas, acetylene, etc. With the advent of World War II, catalytic cracking processes for production of such war materials as aviation gasoline, butylenes, toluene, etc. became important. As a result, many catalytic cracking plants were quickly designed and constructed throughout the United States, and many electrical precipitators were installed in these catalytic cracking plants for recovering fine catalyst from catalyst regenerators. After World War II, the demands of these stragetic materials decreased considerably and the number of yearly installations of precipitators in the area of catalytic cracking has been declining.

Table 21.1 is a summary of the electrostatic precipitator installations for catalyst recovery in fluid catalytic cracking for the period 1940-1962. The number of installations, the total gas volume processed, the efficiency (averaged on a weighted volume basis), and the average precipitator size in terms of gas volume handled are also shown in the table. Figure 21.9 shows the trend in the 5-year average installed precipitator capacity in fluid catalytic cracking operations.

Table 21.2 is a summary of the application of electrostatic precipitators for tar and oil mist removal in the manufacture of carburetted water gas, oil gas, reformed gas, shale and oil gas, and acetylene. The range of collection efficiencies for each application, the inlet gas temperature, and the total gas volume are also given in the table.

Figure 21.10 shows the installed electrostatic precipitator capacity for tar and oil mist removal in carburetted water gas for 5-year intervals between 1940-1960.

Economics. Precipitator cost data for both catalytic cracking application and fuel gas application are summarized in Tables 21.3, 21.4, and 21.5. Figure 21.11 indicates the FOB and erected costs for detarring precipitators. These costs are for a 95% efficient precipitator and are average costs over the period 1959-1969. The costs are flange to flange and do not include ductwork.

Table 21.1

A Summary of Application Data on Electrostatic Precipitator Installations in Catalytic Cracking Units 1940-1967

Yearly Interval	No. of Install.	Total acfm Gas Volume in 1000's	Average Yearly Pptr. Volume over the Period in 1000's	Weighted Eff. on acfm Basis %
1940-1944	24	2,099	420	99.5
1945-1949	13	802	160.4	99.3
1950-1954	3	472	94.4	96
1955-1962	2	488	61	97.2
1966-1967	3	300*	100*	-

*Estimated

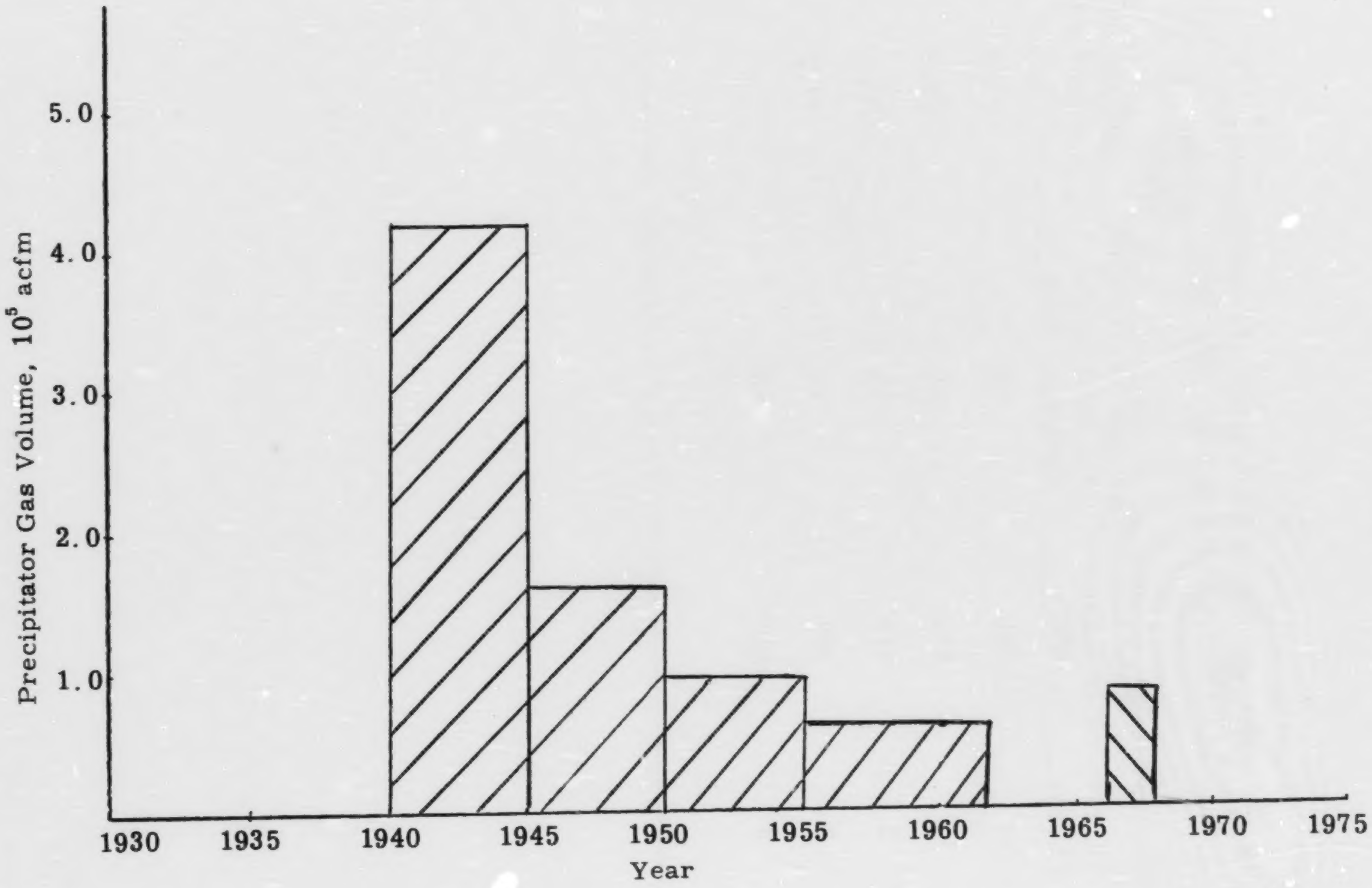

Figure 21.9. Installed Precipitator Gas Volume Trend in Catalytic Cracking Units Over 1940-1962 (Based on Yearly Average acfm per Five Years) and 1966-1967 (Estimated).

Table 21.2

A Summary of Application Data on Electrostatic Precipitator on Various Applications for Removing Tar and Oil Mist 1940-1963

Type of Applications	No. of Install.	Suspended Matters	Total Gas Vol. acfm	Temperature °F	Collecting Eff. %
Carburetted Water Gas	55	Tar and Oil Mist	315,800	70-110	95
Oil Gas	3	Tar and Oil Mist	18,200	80-100	95-98
Reformed Gas	3	Tar and Oil Mist	3,200	80-100	95-98.5
Shale Oil Gas	2	Tar and Oil Mist	20,900	100-200	95-97.5
Acetylene	1	Tar and Oil Mist	42,100	100	99-92

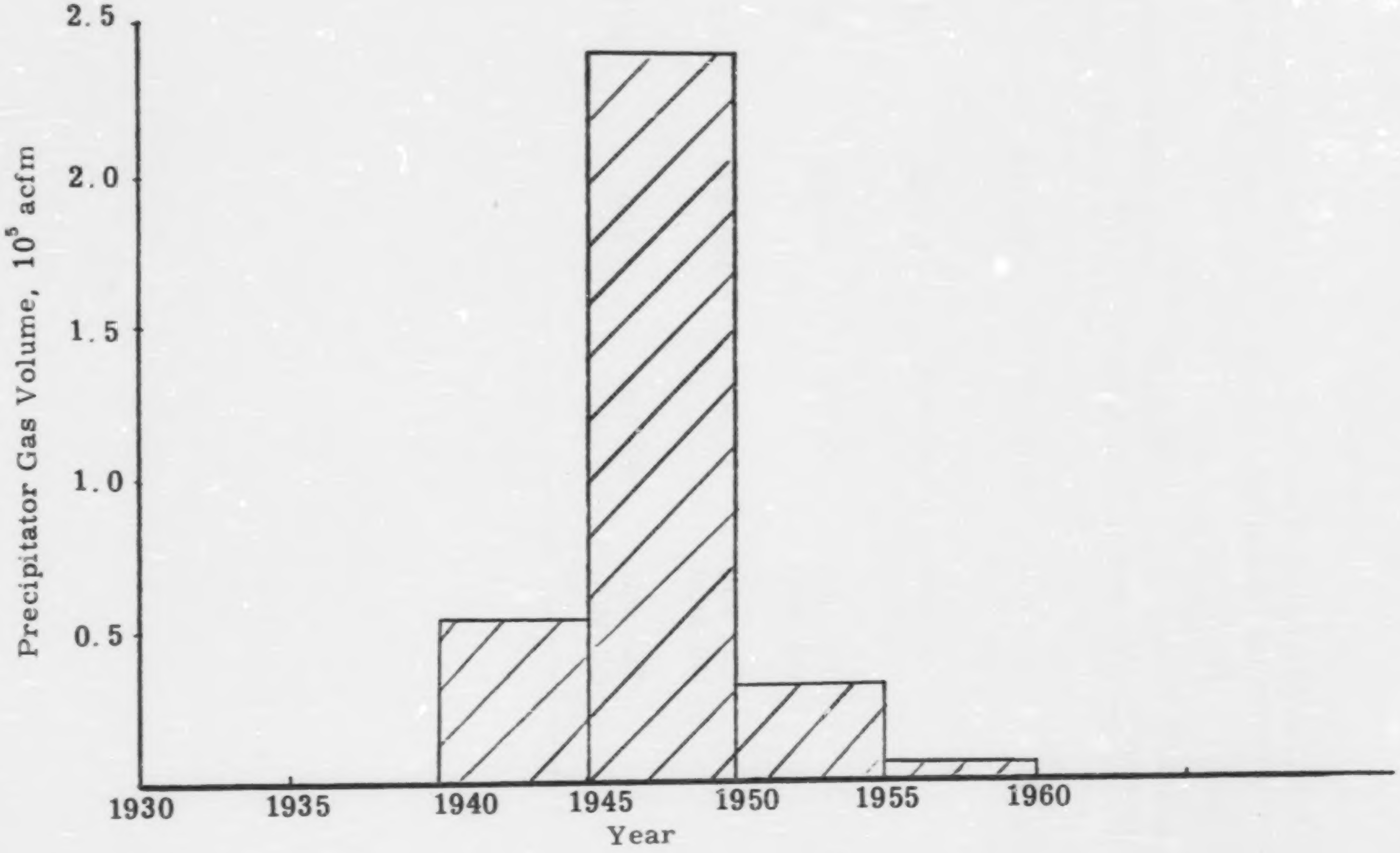

Figure 21.10. Installed Precipitator Gas Volume Trend in Carburetted Water Gas Over 1940-1958 (Based on Total acfm per Five Years).

Table 21.3

FOB Cost of Precipitators for Catalytic Cracking Units 1951-1962

Year	Avg. Gas Volume 1000's acfm *	Avg. Design Efficiency %	Avg. FOB Cost $ / acfm
1951	235 (1)	80	0.426
1953	175 (2)	92	0.78
1954	6 (1)	95	3.78
1958	149.2 (2)	97.5	1.168
1962	254 (1)	96.5	0.48

*Numbers in parenthesis are number of installations used in determining cost.

Table 21.4

Cost* of Detarring Precipitator in Fuel Gas Area 1945-1956

Year	Gas Volume in 1000's acfm	Design Efficiency %	Average FOB Cost $/acfm
1945	34.22	95	1.72
1946	63.6	95	1.94
1947	104.8	95	1.98
1948	40.75	95	2.18
1949	3.2	95	2.25
1950	9.5	95	2.3
1954	6.6	95	2.3
1956	7.2	95	2.3

* Based on the cost of precipitator installation in carburetted water gas.

Table 21.5

Cost* of Detarring Precipitators in Fuel Gas 1959-1969

Year	Gas Volume in 1000's acfm	Design Efficiency %	Cost in $ /acfm	
			Erected	FOB
1959	11.64	95	-	2.44
1960	11	95	4.46	-
1962	23	95	2.63	-
1966	11.4	95	-	2.69
1967	30	95	-	2.45
1968	55.6	95	-	1.1
1969	83	95	1.43	-

* Based on the cost of precipitator installation in coke oven gas.

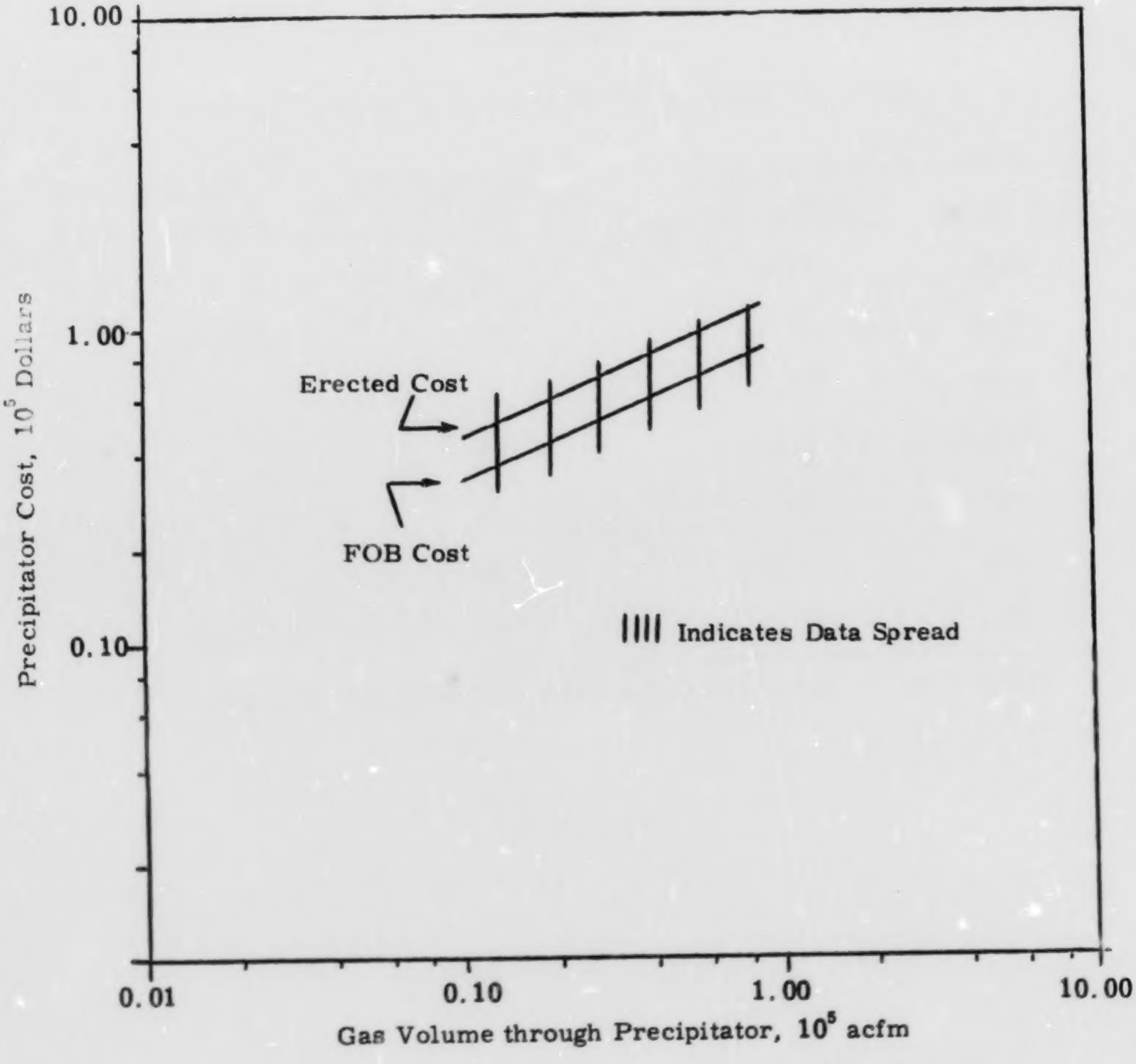

Figure 21.11. Cost of Detarring Precipitators with 95% Collecting Efficiency (1959-1969).

Performance statistics. The important operating parameters in each of the major areas, catalyst recovery and oil mist and tar removal, are summarized in Tables 21.6 and 21.7. The data given in the tables show the maximum, minimum, and average values of gas volume, gas temperature, inlet loading, and collection efficiency.

Design considerations. Selection of the size and power requirements for a particular design specification follows the general methodology outlined in Chapter 9, Part I. The proposed manufacturer selects a precipitation rate parameter for the stated operating conditions. The selection is made from company proprietary data. The required collection electrode area is computed from the Deutsch-Anderson equation. The power supply and sectionalization requirements are established from the proprietary empirical data bank.

Table 21.8 is a summary of design and field test data on two precipitators installed on fluid catalytic cracking units as reported by the American Petroleum Institute.[3] Table 21.9 shows design and performance data on two precipitators collecting powdered catalyst with ammonia injections.

General observations and discussion of trends. The main application of electrostatic precipitation in the petroleum industry has been its use in the collection of catalyst fines from catalytic cracking operations used in the production of high octane gasoline. The surge in use was during World War II when most of the installations were made. With the advent of the jet aircraft in recent years, the demand for high octane aviation gasoline has been replaced by demands for lower octane jet fuel which does not have the same production requirements. Hence, the number of new catalytic cracking units, and the number of electrostatic precipitators for this service has declined as illustrated in Figure 21.3. Electrostatic precipitators as particulate emission control devices are still being utilized on catalytic cracking units, however, and at least 3 units were sold for this designated purpose during the 1966-67 period.[4]

The detarring of fuel gases, although strictly not a part of the petroleum industry, has been included due to close process relationship, and similar precipitator design factors. Again, this application appears to be declining, with uses primarily dictated by a local supply and demand situation, since the use of pipeline natural gas has replaced manufactured gases as the major gaseous fuel source in the United States.

Table 21.6

A Summary of Application Data* for Precipitators in Fluid Catalytic Cracking Application 1951-1962

Parameter	Maximum	Minimum	Average
Gas Volume, 1000's acfm/pptr.	254	6	152.6
Gas Temperature, °F	850	450	610
Inlet Loading, # /hr	2800	77	1444
Collecting Efficiency, %	99.7	80	95
Precipitation Rate,** w ft/sec	0.39	0.1	0.26

* These data based on 7 installations.

** Average of 3 test and 2 design w's.

Table 21.7

A Summary of Performance Data* in Fuel Gas Application 1940-1956

Parameter	Maximum	Minimum	Average
Gas Volume, 1000's acfm/pptr.	16.72	3.5	5.75
Gas Temperature, °F	120	70	90
Inlet Loading, gr/scf**	-	-	0.8-2.5
Collecting Efficiency,	95	95	95
Design Precipitation Rate, w ft/sec	0.55	0.32	0.45

* These data based on 55 installations.

** A tar camera had been applied for efficiency test of detarring precipitators. Testing by this means consists essentially of comparing color densities for filter papers after exposure to the inlet and outlet gas. The average inlet loading in this application is about 0.8-2.5 gr/scf.

Table 21.8

Field Tests on Fluid Catalytic Cracking Units[3]

	No. 1	No. 2
Collecting electrode area, sq ft	. . .	137,760
Rated gas-handling capacity, acfm	73,800 at 515°F	99,900 at 600°F
Test gas-handling capacity, acfm	120,000 at 515°F	138,000 at 470°F
Specified inlet catalyst concentration, gr/scf	5 to 25	4.4 to 22
Test inlet catalyst concentration, gr/scf	25.7	24.4
Guaranteed collection efficiency, %	99.6	99.6
Test collection efficiency, %	99.77	99.77
Test inlet catalyst loading, tons per day	1,085	132
Catalyst	Synthetic	Synthetic
Gas composition, dry basis, mole %:		
CO_2	7.2	9.1
O_2	5.1	3.0
CO	3.2	6.3
N_2	84.5	81.6
Moisture, mole %	34.5	31.4
Ammonia addition rate, lb per hr	15	9.4
Stack catalyst concentration, gr/scf	0.531	0.0538
Stack catalyst loss, tons per day	2.4	0.3
Precipitator operating pressure, psig	0	3.0
Rapping cycles:		
Discharge electrodes	No data	2-hr cycle, 6-sec duration
Collecting electrodes	No data	30-min cycle, 10-sec duration
Computed particle drift velocity, fps	. . .	0.10

Note: Inlet concentration probably is ahead of mechanical collectors.

Table 21.9

Design and Performance Data for Two Powdered Catalyst Precipitators

Rated Gas Volume acfm	100,000	40,000
Gas Temperature	400°F	400°F
Collection Surface Area	34,400	16,300
Design Efficiency	99.6%	99.5%
Test Efficiency	99.7%	99.8%
Test Gas Velocity	-	62,000 at 350°F
Design Precipitation Rate, ft/sec	0.27	0.22
Test Precipitation Rate, ft/sec	0.36[1]	0.39[2]

[1] 1-2 cfm of NH_3 conditioning
[2] 1 cfm of NH_3 conditioning

The major potential use of precipitators in the petroleum industry appears to lie in new applications, two of which are shale oil processing and acetylene manufacturing. However, the total market appears to be small compared to potential applications in other major industries. It should be noted that the use of precipitators for these detarring type applications is not one where air pollution is involved, but instead, one of process and/or product improvement and control.

CHAPTER 21
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3. "Removal of Particulate Matter from Gaseous Wastes," report prepared by the Chemical Engineering Department of the University of Cincinnati for the American Petroleum Institute (1958).

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CHAPTER 22
THE APPLICATION OF ELECTROSTATIC PRECIPITATORS IN THE NONFERROUS METALS INDUSTRY

Applications of electrostatic precipitators in the nonferrous metals industry have been in the removal of particulates for the purpose of air pollution control, cleaning of gas for process purposes, and for recovery of metallic ore-containing dusts for further processing.

The primary nonferrous metals of interest are copper, lead, zinc, and aluminum. The main processes in the extraction of these metals from their ores are: (1) dressing, (2) roasting, (3) sintering, (4) smelting and refining, and (5) electrolytic reduction. These functions are carried out in a number of different types of equipment, depending upon the nature of the ore and the particular plant design.

The commercial use of electrostatic precipitators has been standard practice by copper, lead, and zinc smelters in cleaning the off-gases from the extraction process. Precipitators are also used in the electrolytic reduction of bauxite to produce aluminum.

22.1 HISTORICAL DEVELOPMENT

Although precipitator applications in the smelter field were made initially to abate the smoke nuisance, an even greater usefulness was found in recovering the valuable copper, lead, and zinc oxides and other compounds carried out of the stacks in the form of dust and fume from the furnace operations. Most of the western smelters designed and built their own precipitators under patent license agreements. As a result, striking differences developed in their precipitation practice. Experimental work on the electrostatic treatment of copper converter furnace gases was started as early as 1911. A year later the work was extended to a full-scale unit of 50,000 cfm capacity. As a result of successful experience on converter gases, experimental work was extended to the treatment of gases from lead blast furnaces, roasters, and reverberatory furnaces.

Development of electrostatic precipitators for large smelters was in the direction of large pipes, 12 to 36 inches in diameter and 20 feet long, to handle the large volume gas flows. These large diameter electrodes required rectifier sets to operate at very high voltage to generate the corona and maintain the electric field. Experimental units of up to 48 inches in diameter with rectifiers operating at voltages up to 200 kV were tried on an experimental scale. However, difficulties in handling the high voltage led to the use of plate-type electrodes in a vertical-flow precipitator for this type service.

During the early work on converter gases, it was found that even practically complete removal of the lead fume and clarification of the gas at a temperature of 600-700°F still resulted in a cloudy discharge from the stack. To obtain complete clearance of the stack it was necessary to cool the gases to about 200°F. The explanation lay in the presence of arsenic trioxide and sulfur trioxide in the gas (which were both in the vapor phase at 600-700°F but condensed to fine fumes at 200°F). This suggested the possibility of effecting fractional condensation and separation of the fumes and vapors present by cooling the gases in several steps and using precipitation at each step to collect the component condensed at that temperature. The process was demonstrated by the separation of refined, snow-white arsenic trioxide fume from the furnace gases.

Over the years precipitator applications in smelters have continued to increase. They are used extensively on copper, lead, and zinc reductions; copper roasters, converters, and reverberatory furnaces; sintering machines, blast furnaces, and reverberatory furnaces in lead recovery operations; and roasting of ore concentrates in zinc processing.

The largest smelter precipitator in this country is a central gas-treating installation which was built and placed in operation in 1919. Gas flows of over two million cfm are handled in the precipitator. This precipitator, in addition to the gas volume handled, is notable for the size and capacity of its rectifier equipment which comprises 12 full-wave mechanical rectifiers, each powered by a 100 kV, 75 kVA high-voltage transformer.

In the aluminum industry, the application of electrostatic precipitators was studied as early as 1918 on exhaust gas from baking ovens processing electrodes used in the Hall electrothermic cell for the extraction of aluminum from bauxite. An experimental unit for cleaning ventilating gases from alumina reduction cells was successfully operated in 1951 and resulted in the full-scale installation of precipitators (dry type) to treat over a million cfm of gas.

Subsequently, electrostatic precipitators have been applied to carbon plants, bauxite dryers, alumina calciners, remelt furnaces, etc.

Other miscellaneous applications in the metallurgical field are cadmium recovery from zinc sinter machines, lead blast furnaces, tin smelting, and nickel smelting.

Another application which developed in the early 1920's was the recovery of gold from furnace and chemical operations at the United States Assay Office in New York City. This installation treated 70,000 cfm of gas at an efficiency of over 95 percent. The gold recovered in fume deposits from the precipitator amounted to about $11,000 per annum. Subsequently, similar installations were made at several United States mints. Another application of some importance is the treatment of fumes from furnace operations for the production of metallic silver in connection with the electrolytic refining of copper.

The production statistics in the United States of the four most important nonferrous metals for the period 1935 through 1967 are shown in Figure 22.1.

22.2 NONFERROUS METAL PROCESSING

The process of extracting nonferrous metals from their ores is carried out in a variety of processing equipment. The particular steps in the reduction process and the type of processing equipment vary with the type of ore. The furnaces for roasting, smelting and refining of lead, copper, and zinc are described in the following sections.

Sintering process. A sintering machine is used to convert metallic ores, fines, and plant process dust into larger pieces which can be handled more readily during further plant processing. The sinter machine itself is similar to that used in ferrous metal processing; see Figure 22.2.

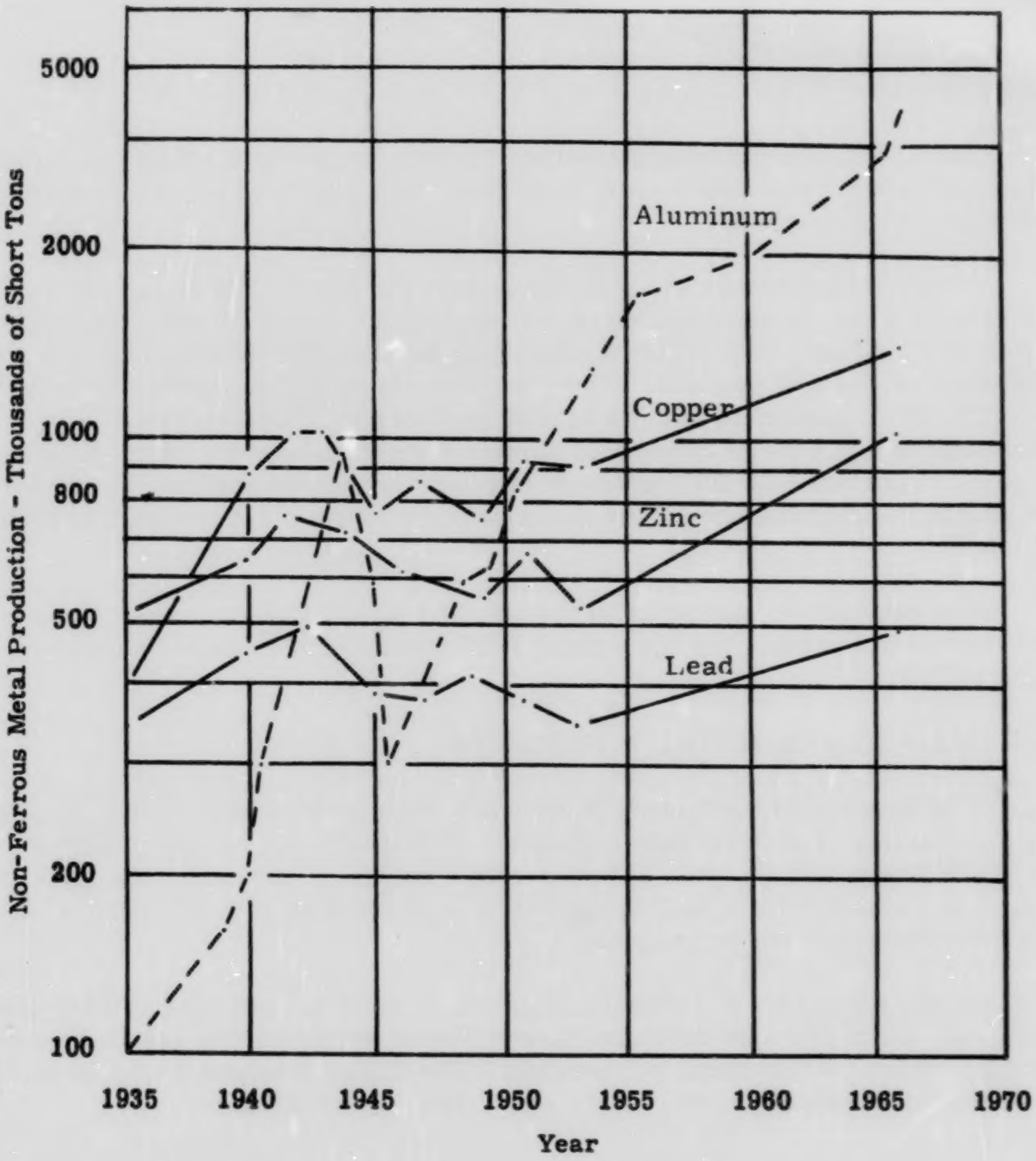

Figure 22.1. Nonferrous Metals Production in the United States.

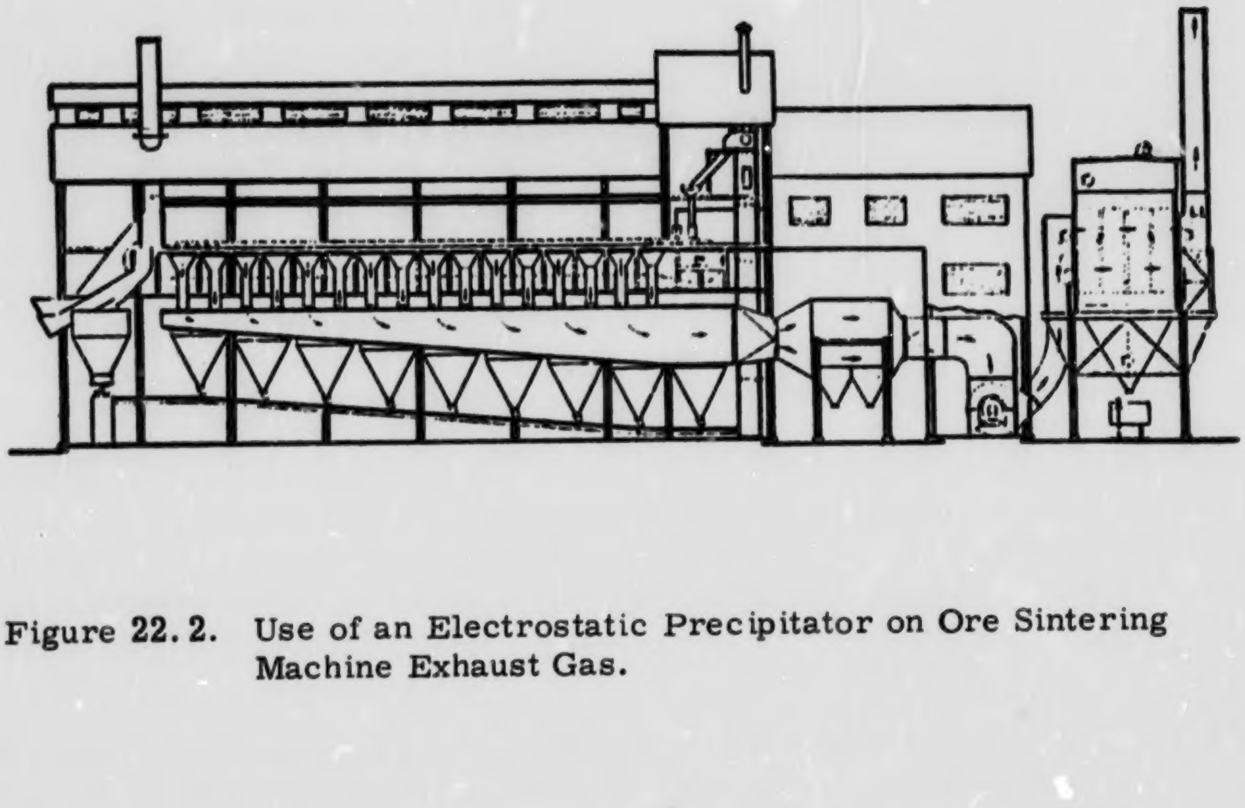

Figure 22.2. Use of an Electrostatic Precipitator on Ore Sintering Machine Exhaust Gas.

In preparing the material for sintering, the ore is mixed with coke and additives, when required. A further preparation, in some cases, consists of dampening the mix so it packs better. The mix is then uniformly spread across the moving grates of the sinter machine and ignited by gas-fired burners. Air is induced through the bed in order to support burning as the grate moves toward the discharge end. At the tail end, the sinter is dropped off the grate, spray-cooled and then transported to the point of further processing.

The combustion air induced down through the bed is collected in a multiplicity of wind boxes located below the traveling grates. From the wind boxes, the gas is transported to the cleaning equipment which usually includes a primary mechanical collector preceding the precipitator. The main fan is located between the precipitator and the stack.

The precipitator usually used in conjunction with the newer zinc sintering machines is a single stage, horizontal flow, plate type. The shell is constructed of steel and insulated by internal guniting if corrosive conditions exist. The hoppers are either trough type, with a screw conveyor along the bottom, or a flat pan type basin with a drag scraper. The discharge electrode system can be of the wire type or any of several discharge electrode configurations described in Chapter 10, Part I. The collecting electrode system is a shielded flat plate type. Discharge electrodes are often cleaned by means of magnetic vibrators, and air vibrators are used on the collecting plates. Gas and/or particulate conditioning, if required to lower particulate resistivity, is by water spray or steam carefully controlled so that the resulting moisture/temperature conditions do not create a corrosive atmosphere.

Ore roasting. Raw ore can be either roasted to completion or partially roasted in preparation for further processing. Prior to roasting, the ore is crushed and screened to control the particle size. Depending on the kind of ore, the type roaster and/or end products, the size can vary from walnut-sized chunks to a fine powder. The nature of the ore to be treated, the kind of roast required, the tonnage of ore to be handled, and the cost of installation and operation, have resulted in a great diversity of forms in the apparatus for roasting ores. The following are the most common types of ore roasters:

Multiple hearth furnace

The multiple hearth furnace is probably the most popular type of ore roaster furnace in use today. It consists of a vertical cylinder of steel plate lined with refractory brick and divided in hearths of arch construction. There can be as few as four hearths or as many as 16 or more depending on the application. The hearth floor and walls, and the roof of the furnace are lined with brick to protect metal parts, as well as to conserve heat. There is a central shaft to which are attached the rabble arms, two for each hearth. This central shaft also contains the pipes which conduct the cooling air or water to the arms. To each arm is fixed seven to nine rakes or rabbles. The rotation of this shaft, which is driven by a motor and train of gears beneath the furnace, is of the order of a revolution in one-half to two minutes. Ore is fed upon the upper hearth, which, being warmed by the heat generated in the roasting operation, serves to dry the crude ore. The rabbles are so adjusted that the ore is gradually moved from the outer edge of the upper hearth toward the center and falls through a drop hole onto Hearth No. 1. The ore then moves across this hearth to a slot near the outer edge, through which it drops onto Hearth No. 2, and thus in zig-zag fashion the ore progresses through the furnace until it finally drops into a car or conveyor beneath the lowest hearth. All furnaces have doors on each hearth for visual observation, repairs, and admission of air. Once the ore is ignited, the reaction itself furnishes enough heat to make the process a self-sustaining one for the most rough roasting. However, some ores are so difficult to roast or so low in sulfur that additional heat must be applied.

Flash roaster

The primary use of flash roasters is in removing sulfur from zinc ore. The sulfur and zinc bearing ore is blown into a combustion chamber along with preheated air. The combustion chamber is usually preheated by gas or oil until the temperature is around 1600 to 1700°F. Ore and heated air are introduced in the combustion chamber and the flash combustion of sulfur occurs. With proper conditions of draft, and with the proper ore composition, the burning becomes self-supporting and the gas jets may be turned off. Under some conditions, such as very high moisture content in the ore, it may become necessary to add auxiliary fuel such as powdered coal or to keep the gas jets on; however, this is unusual and undesirable.

The sinter from the flash chamber drops to the roaster hearths where it is rabbled and further combustion occurs, so that the final sinter usually has less than 1% sulfur as compared to about 30% in the original ore.

Fluid-Bed roaster

A more recent type of roaster is the fluid-bed reactor. Either dry or slurry feeding can be used. Low pressure air is introduced into a wind-box and passes up through an air distribution plate near the bottom of the reactor. Operation of this system is continuous. As feed enters the bed, it is fluidized and immediately brought to roasting temperature. The roasted material overflows into a pipe placed at "surface level" and passes out of the reactor through a cooler.

When a precipitator is used for air pollution control, a flue is connected to the top of the roaster. Prior to entering the precipitator, the gas is tempered in either a dry bottom cooling tower or by mixing with colder atmospheric air. Normally, the type precipitator used to clean the gases coming from roasters is a single-stage, horizontal-flow, plate-type unit. The shell is constructed of steel and the dust hoppers are of pyramidal design. Where condensation may occur, the precipitator is insulated on the outside.

Smelting and refining. Smelting and refining operations are for the purpose of removing unwanted impurity elements and adding alloying materials to produce desired metal composition. The smelting and refining operations are again carried out in a variety of types of furnaces depending on the type of material being processed. The most common types of furnaces are described as follows.

Reverberatory furnace

By definition, a reverberatory furnace is one in which the materials being processed are in direct contact with an incandescent flame. Heating is by direct radiation from the flame, walls and ceiling of the furnace. The furnace is fired by either gas or oil. The primary use of this furnace is for melting solids for the purpose of refining or alloying metals to a specific mixture of materials or assay. This is accomplished by regulating the feed by the additions of chemicals, called "fluxing", or in some cases by the control of temperature as in "sweatbug". The unwanted materials come off as a dross or boil out as gases, such as fluorine, chlorine, etc., which are carried away with the off-gases.

The design of a furnace varies widely, depending on its particular use. Furnace sizes also vary widely from about 500 lb per heat to 50 tons per heat. The smaller units normally have cylindrical shells, and are of the tilting type. The larger furnaces are fixed, having tap-holes for metal pouring. The dross is usually skimmed off. Many large furnaces have double hearths operating at different temperatures for better control. A popular type furnace is one provided with a charging well located adjacent to the furnace but having a common bath. Other furnaces have charging doors at the sides.

Basically, the gas control problem associated with reverberatory type furnaces is one of ventilation. This is because in most cases some form of hooding is involved. The smaller fixed units generally utilize canopy hooding. The tilting types use either adjustable type hoods or plenum hoods. The smallest units often use movable flexible flues. The large fixed units exhaust directly from the furnace with canopy hoods over charging wells and doors. In addition to venting the furnace proper, various operating locations include tap holes, metal runways, holding pots, and pouring areas which are also vented.

The problem of cleaning the gases coming from a reverberatory furnace installation falls into two broad categories. The first is where only particulate is to be removed from the gas. The second is where certain objectional gases, such as produced by the fluxing operation, must be removed in addition to the particulate matter.

A single-stage, vertical-flow, pipe-type precipitator is customarily used if only particulate is to be removed. The shell is constructed of mild steel and is normally uninsulated. The hopper is usually pyramidal in design.

Where gaseous contaminants must be removed in addition to particulates, the precipitator is preceded by a wash tower or scrubber. The contaminant gases are removed by scrubbing with water or a neutralizing solution. The scrubber is usually an integral part of the precipitator. The precipitator proper is usually a single-stage, vertical-flow unit of the wet type, which eliminates the carryover water spray along with the particulates. Because of the corrosive nature of the wet precipitate, the shell is constructed of either steel with a corrosion-proof lining, concrete or wood.

The precipitator in the dry system can utilize weighted wire discharge electrodes or other types as described in Chapter 10, Part I.

Exposed steel pipes are usually used as collecting electrodes in cases where very hot gases are treated by a precipitator. The pipes are left uninsulated, causing gas and dust cooling by radiation. The pipes are between eight inches and twelve inches in diameter, and are arranged vertically between a top and bottom plenum header. Pipe banks are generally arranged for two pass operation. The gas enters the top header section of the first bank, flows downward, reverses and flows upward through the second bank, exiting through the top header.

The cleaning of the exposed pipe collecting electrodes requires vigorous action because of the tacky nature of the dust collected on the cool surface. Usually a form of the swing-hammer pipe rapper is used. The mechanism consists of a motor-driven oscillating shaft running horizontally across the precipitator between rows of collecting pipes. Hammer heads are connected to the shaft by spring leaf arms and strike against anvils attached to the collecting pipes near the bottom. The rapping blow can be varied by adjusting the arc length of the hammer swing and/or controlling the operating time. Vertically operated air vibrators, located on the roof, are employed for discharge electrode cleaning.

The precipitate removal systems used to transport the hopper catch for small and large installations are, respectively, container removal and screw conveyor.

For wet precipitators, discharge electrodes are usually square wires using support insulators with orifices for suspension. In this design, the high voltage insulators are located in compartments on the roof. A bus beam is mounted on the top of the insulators. The discharge electrode assembly is supported from the bus beam by hanger rods which pass through open metal sleeves fastened to the underside of the roof. The sleeves, or orifices, are continuously purged with a small amount of air in order to keep the dirty gas from reaching the insulators in the compartment. Purging is either by inleakage when the system is under vacuum, or through the use of a pressure blower when the system is pressurized.

Since the system is wet and subject to extreme corrosion, a different type of collecting electrode is required. Transite plates are usually used in such wet and low temperature corrosive environments. The plates are constructed from one inch thick transite board, supported at the ends, and spaced on about ten inch centers. A number of standard size sheets, usually eight feet by four feet, are combined to make up the desired size collecting electrode. The plates depend on the wetted surface for electric current conduction and metallic bars placed along the edges of the plates form a grounding network. The precipitator is arranged for vertical gas flow. The transite plates are kept washed by carry-over water entering the precipitator from the preceding scrubber. In addition, sprays are provided under the roof for periodic spraying.

In the wet system, a slanted hopper bottom is generally used, with collected material removal by a slurry hopper system.

Blast furnace

The use of blast furnaces for large nonferrous smelting operations has been largely replaced in recent years by other means. However, of those still in operation, electrostatic precipitators are used more often with the larger blast furnaces, which are nearly all in the copper and lead smelting process. Many metals such as cobalt, nickel, tin and silver are by-products of the reduction operation of copper and/or lead smelting.

In the copper smelting process, the blast furnace is a shaft furnace. The charge consists of ore, fuel, and flux. The hot gas from the combustion of the fuel rises up through the furnace countercurrent to the descending charge. As reduction in slagging proceeds, the matte and slag formed descend to the crucible. Smelting in a blast furnace may be performed by the reducing, pyritic, or semi-pyritic process.

In the reducing process, the oxide or rough-roasted ore is mixed with flux and only enough carbonaceous fuel to furnish the necessary heat to carry out the reduction of the oxide. The blast of air should oxidize only the carbon, and none of the sulfur.

In pyritic smelting, an attempt is made to save coke by using the heat of the oxidization of sulfur to promote the reactions of matting and slagging. Although theoretically there may be sufficient heat generated in the furnace for this purpose, it is usually necessary to use 1-3% of coke.

The most common type of smelting today is the semi-pyritic process, because it is carried out with ores which do not contain enough sulfides for pyritic smelting and yet may not require roasting. The amount of coke used varies from 3-16% of the weight of the charge.

A blast furnace used to smelt lead ores is slightly different from that used for copper reduction. The difference is in the construction of the crucible. In the lead blast furnace the crucible is deep and the lead, as it is reduced in the furnace and collects in the crucible, is removed from beneath the slag and matte by means of a lead well or a syphon pump.

A nonferrous blast furnace is connected to the inlet of an electrostatic precipitator through a duct system. As a temperature control measure, provisions are made either to dilute the flue gas with ambient air, or to cool it with a dry bottom water spray gas cooler.

The type of precipitator used in connection with a nonferrous blast furnace is a single-stage, two-pass, vertical-flow pipe unit similar to that used on reverberatory furnace gas. The shell and pyramidal hoppers are fabricated of mild steel and are usually left uninsulated.

Converters. In the nonferrous metallurgical industry, a converter is used to convert matte (a mixture of iron and copper sulfide) to metallic copper. In addition to the base metal produced in the converter, some precious metals such as gold and silver are recovered. Most of the impurities such as arsenic, antimony, lead, and zinc are volatilized as oxides and are carried out of the converter with the sulfur dioxide exhaust gas.

During the conversion process, air in small bubbles is forced through the molten matte contained in the converter. When the blast of air from the tuyere mouth enters the molten matte, the sulfides are oxidized. The sulfur dioxide formed escapes through the converter mouth while the ferrous oxide unites with the silica of the flux to form a slag. The heat of formation of this slag, together with that produced in the oxidation of the sulfur and iron, is sufficient to keep the bath molten.

There are several kinds of converters in general use. The two most popular are the "Bessemer Converter" and the "Tilting Horizontal Converter." The gas discharge ports of the converter are coupled to a flue system which transports the gas to an electrostatic precipitator, where the condensed metallic fumes are removed. The gas is cleaned preparatory to processing in the acid plant.

The precipitator employed for the removal of metallic fume from the gas emitted by a nonferrous metal converter is of the single stage type. The gas flows horizontally through ducts formed by the collecting electrodes. The dust hoppers are of the pyramidal type. The shell and hoppers are connected to their respective flue systems by nozzle type transitions. The discharge electrodes are weighted twisted squares suspended from support insulators with bushings. The collecting electrode plates are made of standard corrugated steel sheets. The dust precipitated in the troughs is shielded from the main gas stream, thus minimizing reentrainment. The plates are ten to twenty feet in height and from three to six feet in the direction of gas flow.

Cupola furnace. A cupola in the nonferrous industry is used as a furnace to melt and reduce copper, brasses, bronzes, and lead. It is essentially an open top refractory lined cylinder, equipped with a charging door about one-half of the way up, and air ports (known as tuyeres) at the bottom. Air is supplied from a forced-draft blower. Alternate charges of metal, coke and limestone and/or flux are placed on top of the burning coke bed to fill the cupola. The heat generated melts the metal, which is drawn off through a tap hole.

When a precipitator is used for air pollution control, a flue is connected to the top of the cupola. A by-pass hood or damper is installed as part of the flue connection. Prior to entering the precipitator, the gas is cooled in either a dry bottom spray cooling tower or by mixing with cooler atmospheric air.

The typical precipitator used to clean the gas coming from a cupola is a single stage, horizontal flow, plate type unit. The shell is constructed of steel and is operated without thermal insulation.

22.3 ELECTROLYTIC REDUCTION OF ALUMINUM

Metallic aluminum is produced by electrolytic reduction of alumina (Al_2O_3) dissolved in a molten bath or cryolite ($AlF_3 \cdot 2NaF$). The electrodes used are either prebaked carbon or a continuous anode type. Both electrodes are composed of petroleum coke with pitch as a binder. In the prebaked operation the anodes are prefabricated and baked prior to use. In the Soderberg cells, a paste mixture of coke and tar is added to a compartment over the cell and baked by the heat into a solid anode as it descends. Electrolytic, thermal, and chemical action in the cell evolves carbon and alumina dust, and both particulates and gaseous fluorides. Present practice is to incorporate an effective hood over and/or around each "pot" through which sufficient ventilating air is drawn to contain the emitted fume, ore and carbon dust as well as the pitch in the case of Soderberg electrodes. The gas from a number of pots is collected in a common flue system and then passed from a mechanical collector which removes about 50% of the solids. The collector, usually a mechanical type, discharges the gas into the precipitator where most of the remaining fluorine particulate is removed. Before the gas is discharged to the atmosphere, it is passed through a scrubbing tower to remove the residual gaseous fluorine. Figure 22.3 is a schematic of an emission control system installed for cleaning Soderberg cell gases at an aluminum plant in Germany.

The gas volume handled is about 1500 to 2000 acfm per pot. The temperature is about 200° F, while the moisture content is low, usually less than 2.5% by volume. The moisture is mostly atmospheric moisture carried in ventilating air and can be as low as 1%. See Figure 22.4.

The precipitators used in connection with the electrolytic reduction of alumina are single-stage, horizontal-flow duct units. The hoppers are of a pyramidal design. Because of the relatively low temperature and humidity, no heat insulation is used.

Table 22.1 shows design and performance data for two precipitators installed on aluminum reduction furnaces.

22.4 PRODUCTION OF PRIMARY COPPER

Production processes. Primary copper is produced by smelting or leaching methods, usually followed by electrolytic refining. In general, sulfide ores are treated by smelting, and oxide ores are treated by leaching.

In the smelting process there are usually five steps. These are:

(1) concentration by selective flotation,

(2) roasting,

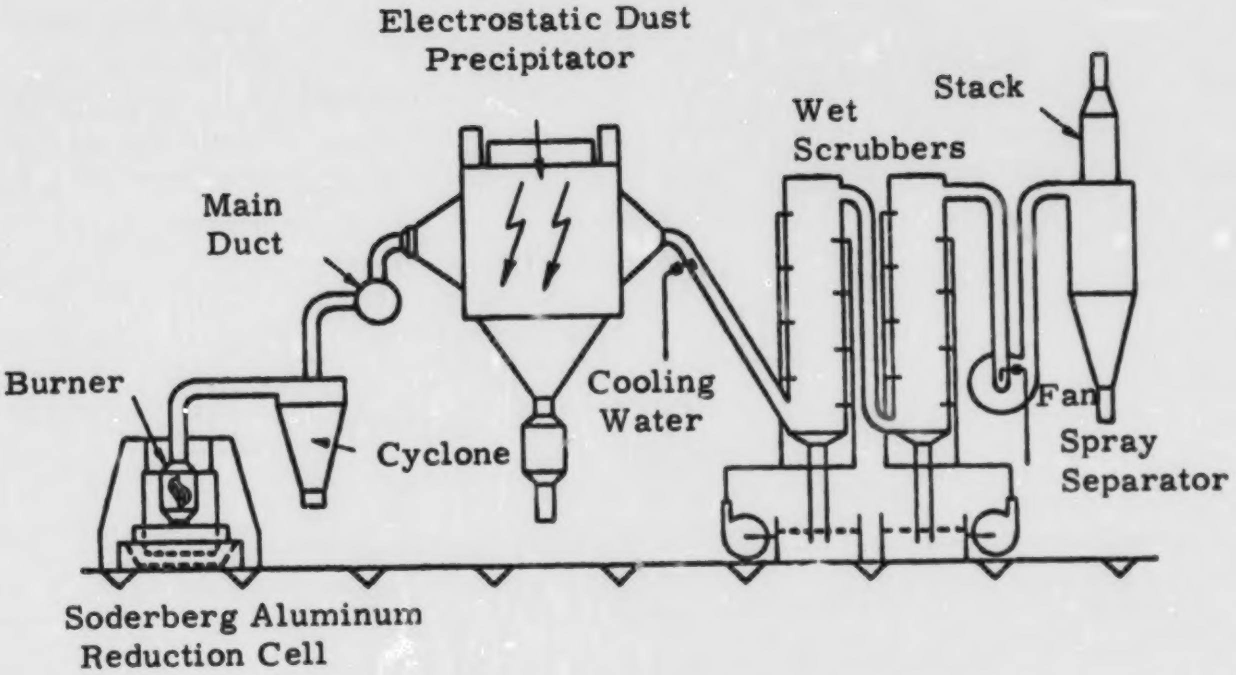
Electrostatic Dust
Precipitator
Stack
Wet
Scrubbers
Main
Duct
Burner
Cooling
Water
Cyclone
Fan
Spray
Separator
Soderberg Aluminum
Reduction Cell

Figure 22.3. Purification System for Soderberg Cell Gases.

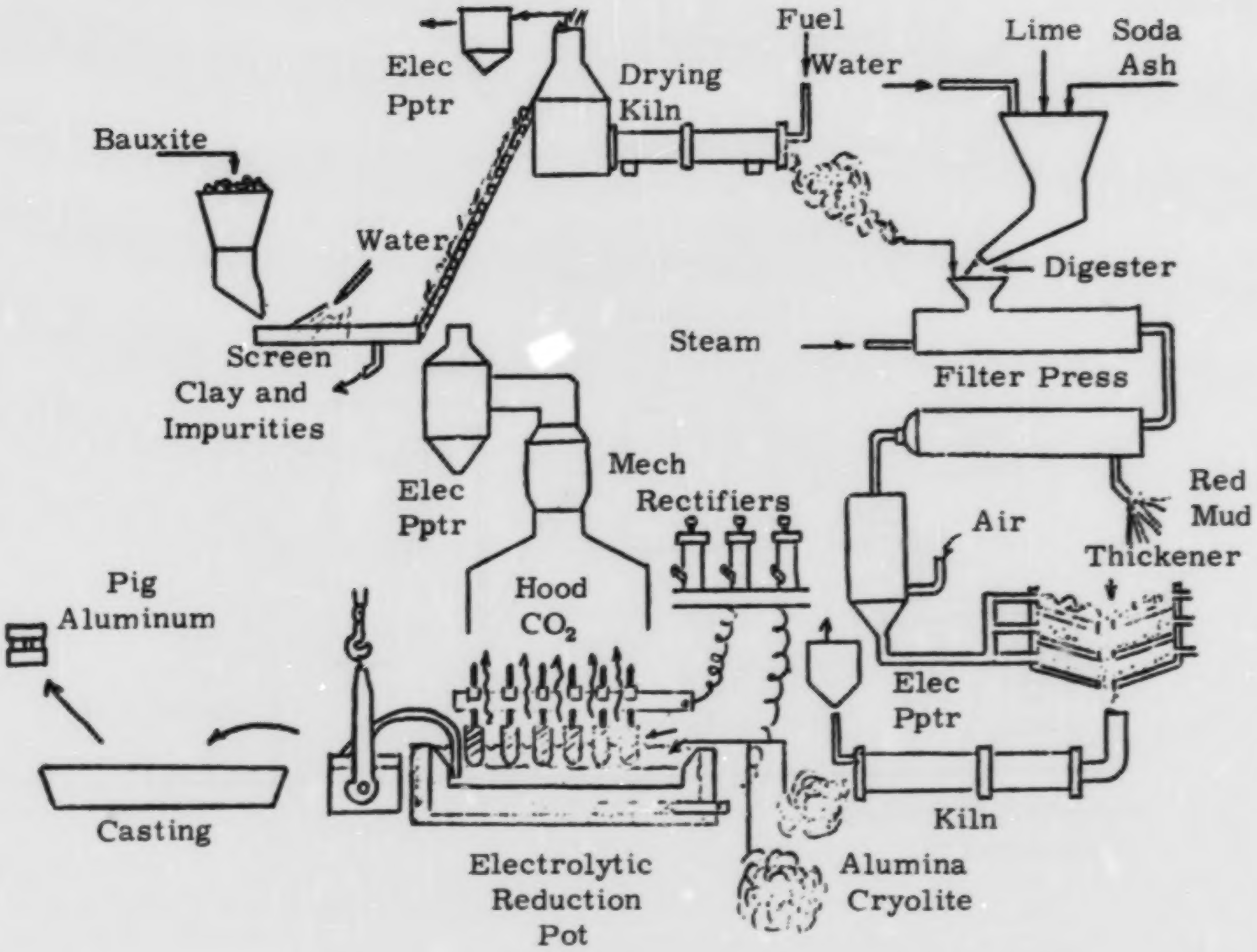
Fuel
Water
Lime
Soda Ash
Elec Pptr
Drying Kiln
Bauxite
Water
Digester
Steam
Filter Press
Screen
Clay and Impurities
Elec Pptr
Mech Rectifiers
Red Mud
Air
Thickener
Pig Aluminum
Hood
CO_2
Elec Pptr
Kiln
Casting
Electrolytic Reduction Pot
Alumina Cryolite

Figure 22.4. Flow Diagram for Aluminum Production.

Table 22.1
Design Performance Data for
Aluminum Reduction Furnace Precipitators

	No. 1	No. 2
Rated gas flow, acfm	57,500	115,000
Gas temperature, °F	212	212
Collecting surface area, sq ft	8,820	17,700
Gas velocity, ft/sec	3.93	3.9
Design efficiency, %	90	90
Design precipitation rate, ft/sec	0.25	0.25
Test gas volume, acfm	55,000	115,000
Test efficiency, %	98	91.5
Test precipitation rate, ft/sec	0.41	0.27
Corona power, w/1000 cfm	180	110[1]

[1] Moisture content lower than precipitator No. 1 (0.9 as compared with 2.4)

(3) smelting,

(4) converting,

(5) fire, or electrolytic refining.

All of these steps, with the exception of electrolytic refining, may result in the generation of particulate matter.

Selective flotation is almost exclusively used in beneficiation of sulfide ores. In flotation processes, the finely ground ore is mixed with water and selected chemical reagents and agitated with air to produce a heavy froth. The reagents are selected such that the mineral particles are carried to the surface in the froth while gangue particles remain in suspension and are pumped to a settling lake. The resulting concentrate contains copper sulfides as well as some sulfides of other metals. After drying, this concentrate is ready to be roasted, or in some cases, sent directly to the smelter.

The objectives of roasting are: (1) to eliminate some of the sulfur from the concentrate by changing the sulfides into gaseous oxides and (2) to volatize zinc, arsenic, and antimony present as impurities.

Roasting is conducted in multiple hearth furnaces or in fluid bed roasters as described previously. After removing some of the sulfur as SO_2, the roasted material is known as calcine and is ready for smelting.

The exit gas temperature during roasting is about 1200°F, and gas volumes range from 5,000 to 1,500,000 cu ft, depending on size of the roasters, with an SO_2 concentration of 5-14%. The SO_2 may be recovered and used to produce sulfuric acid. The dust in the stack gases ranges from 3 to 6% of the roaster feed and produces a dust concentration of 6-10 grains per std cu ft. The dust is primarily composed of sulfide concentrate feed materials. Gas cleaning may be accomplished by wet scrubbing, bag houses, or electrostatic precipitators.[1,2] Some operating statistics of multiple hearth roasters are given in Table 22.2.[1,2]

Stack gas volumes from fluid bed roasters average from 6,000 to 10,000 scfm, with SO_2 concentrations of 8-14%. Dust recovery is affected by cyclones and electrostatic precipitators. Table 22.3 presents some operating characteristics of fluid bed roasters.

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[1] Refer to the bibliography for this chapter.

Table 22.2

Operating Data of Multiple Hearth Roasting Furnaces

Furnace	
Utilization	roasting of mixed copper concentrate
Capacity	140 ton/day
Charge	mixed copper concentrates and flux
Moisture Content	8%
Waste Gas Factors	
Off Gas Volume	4500-5000 scfm
SO_2 - Content	5-14%-vol
Dust	
Content in Raw Gas	6-10 grains/scf
Chemical Composition	9% Cu, 10% S, 26% Fe
Dust Removal	
Type of Separator	ESP and scrub
Efficiency Rate	95-99%

Table 22.3

Fluid Bed Roasters

Furnace	
Utilization	Roasting of mixed cu concentrates
Capacity	140-250 tons/day
Charge	Mixed copper concentrates
Waste Gas Factors	
Off Gas Volume	6,000-10,000 scfm
SO_2 - Content	8-14%-vol
Dust	
Dust Removal Types of Separator	Cyclone in series, ESP
Efficiency	95-99%

After roasting, the first step in smelting is to produce a molten sulfide of iron and copper. This is generally accomplished in reverberatory furnaces ranging from 22-38 ft in width and 96-125 ft in length. The furnace is fired by burners located at one end and the charge is fed in along the furnace walls. The slag formed during melting is lighter than the matte and easily separates from it. When needed, the molten matte is drawn from the furnace through a tap hole near the bottom of the furnace.

Stack gas volumes from matte furnaces average 45, 000 to 65, 000 scfm with SO_2 concentrations of 0. 5 to 3. 5%. The dust load averages 2 to 5 gr/scf, with the dust being composed primarily of copper, zinc, silica, and sulfur.[1, 2] The dust is usually greater than 5μ in diameter, with only limited amounts of smaller particulate, and collection equipment usually consists of electrostatic precipitators or baghouses. Some operating data on copper reverberatory furnaces are given in Table 22. 4.[2]

Matte converting

Copper has a lower affinity for oxygen than iron has for sulfur. This, coupled with the fact that oxidation of iron and sulfur liberates large amounts of heat, makes it possible to blow copper matte with air to oxidize any iron sulfide present and form SO_2 and iron oxide. If any copper is oxidized, it immediately reacts with iron sulfide to produce iron oxide. Since iron oxide forms an easily fusible slag with silica, it is necessary to add sufficient silica to the molten mass to form the desired slag, and all the iron can be removed as liquid iron silicate.

These reactions are usually carried out in a copper converter consisting of a refractory-lined horizontal drum about 13 ft in diameter and approximately 30 ft long. This drum is fitted with an opening at the top through which the furnace is charged and through which gases escape when the furnace is in operation. Along one side of the furnace is a large pipe which delivers air to the furnace through a series of small pipes called tuyeres that enter the furnace through the side.

Matte is charged into the converter directly from the smelting reverberatory furnace and the blow begins immediately. Silica is added as a flux as the temperature of the bath rises and slag is removed and ultimately the furnace is nearly full of molten copper sulfide. At this point slagging is complete and the final blow begins. Oxidation continues until all the

Table 22.4

Operating Data for Reverberatory Furnaces

Furnace	
Utilization	Melting of copper concentrates
Capacity	500 ton/day
Waste-Heat Steam Production	300 ton/day
Charge	Hot roasted material (500° C), quartz and limestone, molten copper slag
Waste Gas Factors	
Off Gas Volume	45,000-65,000 scfm
CO_2-Content	13%-vol
Dust	
Content in Raw Gas	2-5 grains/scf
Chemical Composition	6.2% Cu, 13.0% Zn, 13.6% S
Dust Removal	
Type of Separator	Wet scrubbers and electrostatic precipitator
Efficiency	95-99%

sulfur has been removed and molten copper remains. If the metal is cast and solidified at this time, the product is known as blister copper because of gases liberated during solidification which produce blisters on the surface of the cast metal.

The chemical reactions occurring in the converter are as follows:

$$FeS + \frac{3}{2}O_2 = FeO + SO_2 \text{ (gas)}$$

$$FeO + SiO_2 \rightarrow FeO \cdot SiO_2 \text{ (slag)}$$

$$Cu_2S + O_2 = 2Cu + SO_2 \text{ (gas)}$$

These reactions are exothermic and continue as long as there is sulfur in the metal and as long as oxygen is supplied. The operating temperature is usually about 2250° F. Off-gas volumes of 12,000-15,000 cfm are common and higher flow rates of over 30,000 scfm have been reported. The SO_2 content of the off-gas usually averages about 3.5 to 7.0%. The dust level in the off-gas runs from 5 to 6 grains per std cu ft and consists of copper, zinc, silica, sulfur, and iron. Dust removal from the effluent stream is usually accomplished by means of electrostatic precipitators or baghouses.[3] Some operating data are presented in Table 22.5.[3]

Refining of blister copper. Blister copper is purified either by fire or electrolytic methods. The furnaces used for fire refining are small reverberatory type furnaces or a revolving furnace similar to the copper converter. There are a group of fuel burners to melt the charge or maintain it in the molten condition throughout the refining operation. Air is forced through the molten material to secure complete oxidation of all impurities and then the oxides are allowed to rise to the surface of the quiet pool, from which they are skimmed and then returned to the converter. The oxidizing treatment is followed by a reducing treatment known as poling. This is done by forcing the ends of green logs into the pool of molten metal. The highly reducing gases resulting from the destructive distillation of the green logs reduce most of the copper oxide present in the metal. The resulting product is called tough-pitch copper and is a useable product in this form. It will, however, contain any gold or silver which may have been present in the copper. No information is available on emissions produced during poling of the copper.

Table 22.5

Operating Data for a Copper-Matte Converter

Furnace	
Utilization	Converting of copper matte
Capacity (copper)	35 ton/charge
Specific Furnace Load	2 charge/day
Heat Consumption	Exothermic process
Charge	Molten copper, matte and solid quartz
Moisture Content	None
Waste Gas Factors	
Off-Gas Volume	12,000-13,000 scfm
SO_2-Content	4%-vol
Dust	
Content in Raw Gas	5-6 grains/ft^3
Chemical Composition	1.2% Cu, 18.0% Zn, 10% S
Dust Removal	
Type of Separator	Electrostatic and cloth
Efficiency	95-99%

A second method of purifying copper is electrolytic refining. When an impure copper anode and a cathode are immersed in a solution of copper sulfate and an electric current is imposed across the cell, copper from the anode will enter the solution, and copper ions in the solution will migrate to the cathode and be deposited as metallic copper. Impurities from the anode are liberated and settle to the bottom of the cell as a slime. Pure copper is precipitated on the cathode. No particulate emissions are evolved in this process.

Figure 22.5 is a flow diagram showing the steps in the copper extraction and refining process.

22.5 PRODUCTION OF PRIMARY LEAD

Oxide and sulfide types of lead ore are the only ones that occur in sufficient quantities to be commercially important. Oxide ores can be directly reduced in a lead blast furnace charged with ore, coke and a flux. Sulfide ores must first be converted to oxides. This is accomplished by roasting or sintering in an oxidizing atmosphere on a traveling bed sinter machine to combust the sulfur and produce a sintered product. After sintering, the material is charged into a lead blast furnace for reduction to lead bullion.

Emissions from sinter beds. Sinter beds used in sintering of lead sulfide ores range in size from 3.5 ft wide and 22 ft long to 10 ft wide and about 120 ft long. The sinter bed travels in a continuous loop and capacities range from about 1.5 to 2.75 tons of charged material per square ft of bed per day.[1]

The predominate chemical reaction occurring in the sinter bed is the oxidation of lead and other metal sulfides present. In order to achieve good combustion efficiencies, the sulfide ores are first finely ground and pelletized with coke breeze. The green pellets are placed on the sinter bed, ignited, and combustion is sustained by forced air flow through the bed. Stack gas volumes range from 100 to 220 scfm/sq ft of bed and may contain up to about 8% SO_2. Gases also contain small amounts of water vapor, carbon dioxide, hydrogen fluoride and silicon tetrafluoride as atmospheric contaminants.[1]

The dust carried by the gases from the sinter bed may amount to as much as 5-20% of the feed to the sinter machine. The dust usually contains 40-65% lead, 10-20% zinc, and 8-12% sulfur with traces of other elements

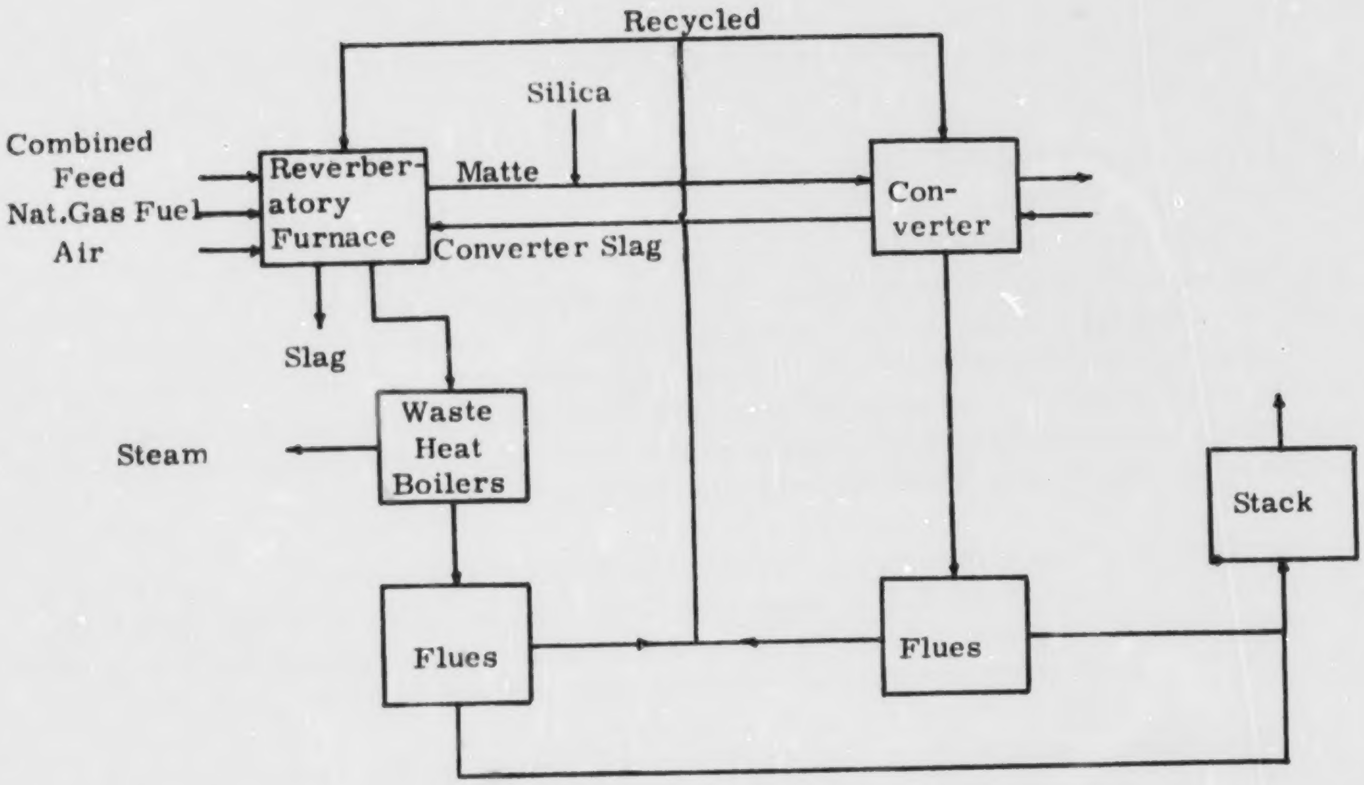
Recycled
Silica
Combined Feed
Nat.Gas Fuel
Air
Reverberatory Furnace
Matte
Converter Slag
Con-verter
Slag
Waste Heat Boilers
Steam
Flues
Flues
Stack

Figure 22.5. Flow Diagram Showing Steps in the Copper Extraction and Refining Process.

including selenium, arsenic, cadmium, and tellurium. The amount of each will generally correspond to the composition of the sinter bed material.[1,4] Table 22.6 presents some general operating data on sintering installations.[1,4]

Because of the high metal content of the particulate of the off-gases, it is economically justifiable to collect them. Baghouses and electrostatic precipitators with operating efficiencies of 95-99% are commonly used. The particulate matter in the gas stream is generally greater than 5μ in size, although there may be some fine metallic fumes.[4] After collection, the particulate matter is recirculated into the sinter bed with new feed ore.

Lead blast furnace operation. After conversion of the sulfides on the sinter machine, the material is charged into a blast furnace along with coke and limestone. In the blast furnace, the following reactions occur:

$$C + O_2 \rightarrow CO_2$$

$$CO_2 + C + \text{heat} \rightarrow 2CO$$

$$PbO + CO + \text{heat} \rightarrow Pb + CO_2$$

The molten lead collects in the furnace hearth along with the iron-lime-silicate slag. Because of the marked difference in density between the lead bullion and the slag, separation is easily effected.

Some general operating data for blast furnaces are presented in Table 22.7. Blast furnace capacity ranges from 60 to 500 tons per day throughout. From 8 to 13% of the charge will be coke and from 240-800 scf of flue gases are produced per lb of coke burned. The dust content of the raw stack gases ranges from 2 to 6.5 grains per scf. Primary cleaning of the raw gas may be done by centrifugal separators, but high cleaning efficiency usually requires the use of baghouses or electrostatic precipitators.[4]

Refining of lead. The end product of the lead blast furnace is an impure lead which may contain copper, nickel, arsenic, silver, gold, tin, antimony, and other minor constituents. In order to purify the lead and remove the other elements, further refining is necessary.

In electrolytic refining, the lead bullion is cast into anode plates and placed in electrolysis tanks. In this operation, electric current causes the

Table 22.6

Operating Data for Sintering Installations

Sinter Machine		
Utilization	Roasting and sintering of galena	
Capacity	50-250 tons/day galena	
Feed	Concentrate, pre-roasted ore, flux	
Moisture content of charge	6-10%	
Ignited by	Gas, oil, coal dust	
Waste-Gas Factors		
Volume		
Sulphur dioxide content	48-56 ft^3/lb sinter	
Temperature prior to dust removal	1.5-8% by volume 300°C	
Dust		
Dust content of raw gas	0.9-6.5 grains/scf	
Chemical composition	40-65% lead 10-20% zinc 8-12% sulphur and traces of elements corresponding to composition of mixture	
Dust Precipitation		
Equipment	Primary:	Secondary:
	centrifugal separators	electrostatic precipitator, bag filters
Efficiency	80-90%	95-99%

Table 22. 7

Operating Data for Blast Furnaces

Reducing Furnaces		
Utilization	Reduction-smelting of sinter, ore, lead scrap and slag with coke and flux	
Capacity	60-500 ton/day throughput	
Fuel Consumption	8-13% coke	
Waste-Gas Factors		
Volume	240-800 scf/lb coke	
CO_2 - Content	Less than 16% volume	
Dust		
Dust Content of Raw Gas	2-6.5 grains/scf	
Chemical Composition	Greatly varying shares of sulphates, oxides, lead sulfide and coke dust	
Dust Precipitation		
Equipment	Primary:	Secondary:
	Centrifugal Separators	Electrostatic precipitators, bag filters
Efficiency	80-90%	95-99%

lead to dissolve in a complex water solution at the anode, and at the same time lead already dissolved is plated out on the cathode in pure form. The impurities are left behind at the anode or remain in solution and must be reclaimed in subsequent operations. No particulate emissions are expected during purification of lead by electrolysis.[4]

In kettle or reverberatory furnace refining of lead bullion, the treatment is more complex, but the recovery of the by-products is simpler and faster. There are several such refining processes or modifications, but basically they have the following steps:

(1) Softening, where oxygen is employed to produce a liquid slag containing the more readily oxidizable elements, including antimony, arsenic, tin, and zinc.[4]

(2) Noble element removal, where zinc is added to produce a dross containing the silver, gold, copper, and nickel,[4] and

(3) De-bismuthing, where magnesium or clacium is added to form a dross which can be skimmed off.[4]

Each of these dross materials must be further treated to obtain separation of the desired elements. The remaining lead usually has a purity of 99.999%.

Very little information exists on atmospheric emissions produced during lead refinement. Undoubtedly, the reverberatory stack gases contain small amounts of lead and zinc, but quantitative estimates are not available.

A final method of purifying lead is the Imperial process, which is basically a tower distillation method. In this process, lead bullion from a blast furnace is heated in a vacuum chamber to evaporate lead, zinc, and other volatile metals. These elements may then be separated in the vapor phase by controlling the temperature along the length of the condensation column. No particulate emissions are produced in this process. Figure 22.6 is a schematic diagram of the process showing the location of electrostatic precipitators for lead blast furnace cleaning. Figure 22.7 shows a rotary furnace rather than a blast furnace.

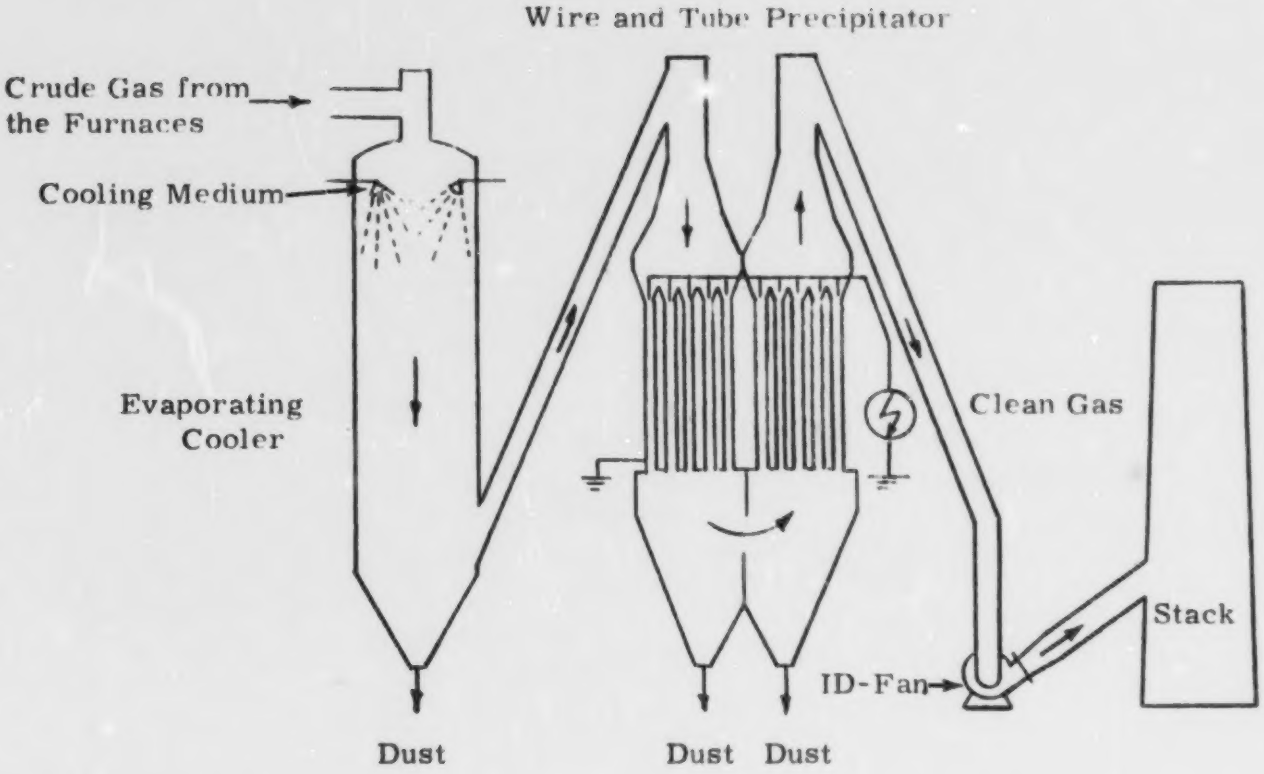
Wire and Tube Precipitator
Crude Gas from the Furnaces
Cooling Medium
Evaporating Cooler
Clean Gas
Stack
ID-Fan
Dust
Dust
Dust

Figure 22.6. Schematic Diagram Showing Electrostatic Precipitators Used for Blast Furnace Gas.

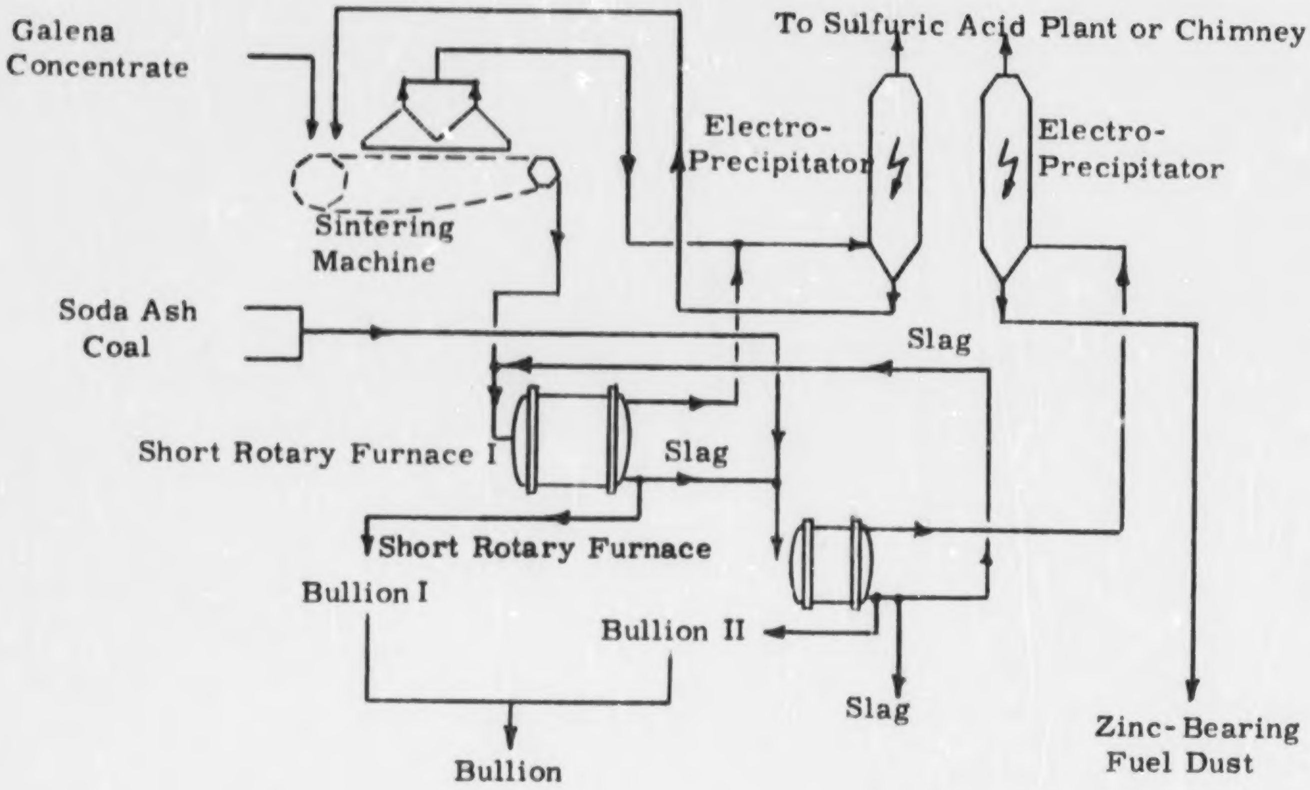
Galena Concentrate
To Sulfuric Acid Plant or Chimney
Electro-Precipitator
Electro-Precipitator
Sintering Machine
Soda Ash
Coal
Slag
Short Rotary Furnace I
Slag
Short Rotary Furnace
Bullion I
Bullion II
Slag
Bullion
Zinc-Bearing Fuel Dust

Figure 22.7. Electrostatic Precipitator Operating on a Rotary Furnace.

22.6 ZINC REDUCTION

Most zinc produced in the United States is extracted from ores containing both zinc and lead sulfides, although some zinc and copper-zinc ores are processed. The ores are concentrated, usually by flotation, prior to the extraction process.

The concentrated ore is dried and processed in a roaster to convert the zinc sulfides into a dense oxide. The sulfur dioxide off-gases are often used to produce sulfuric acid by the contact process, which is described in the chapter on chemical industry applications.

Metallic zinc is produced from the roasted ore by retorting, electrolysis, or fractional distillation.

Roasting of zinc ores takes place in one of several types of furnaces. In this country, there are 12 plants reportedly roasting zinc ores using fluid bed, flash, multiple hearth, and Ropp roasters. The Ropp roaster is the oldest type furnace in use. It is a type of reverberatory furnace which has largely been replaced by more modern roasters. Table 22.8 lists the operating temperature, capacity, dust emission, and percentage of SO_2 in the off gas.

Figure 22.8 is a flow diagram for a zinc plant utilizing a retort process for zinc reduction. In this process the roasted ore is mixed with coal and briquetted prior to charging into the retort. As indicated in the flow chart, the roaster calcine is sintered to agglomerate it prior to subsequent processing. Table 22.9 shows the capacity, dust, and SO_2 concentration in the off gas for typical zinc sintering operations.

In the retort process, the charge, consisting of roasted ore and coke, is charged into the retort and heated to a temperature where the zinc oxide reacts with the carbon to produce zinc metal and CO_2. At these temperatures, the zinc comes off as vapor and is led to a condenser, where condensation to molten zinc occurs. The molten zinc from the condenser is tapped at intervals and cast into billets.

In the electrolytic process, sulfuric acid is used to dissolve the zinc ore to form the electrolyte for the electro-deposition process. Zinc is plated out on aluminum electrodes, which are then stripped for recovery of the zinc.

Table 22.8

Typical Zinc Roasting Operations[1]

Type of Roaster	Operating Temp, °F	Feed Capacity ton/day	Dust in Off Gas % of feed	Off Gas SO_2	Volume scfm
Multihearth	1200-1350	50-120	5-15	4.5-6.5	5,000-6,000
Multihearth[2]	1600-1650	250	5-15	4.5-6.5	5,000-6,000
Ropp[3]	1200	40-50	5	0.7-1.0	20,000-35,000
Fluid Bed[4] (Dorr-Oliver)	1640	140-225	70-80	7-8	6,000-10,000
Fluid Bed[2] (Dorr-Oliver)	1650	240-350	75-85	10-12	6,000-10,000
Fluid Bed (Lurgi)	1700	240	50	9-10	6,000-10,000
Suspension	1800	120-350	50	8-12	10,000-15,000
Fluid Column	1900	225	17-18	11-12	N. A.

---

[1] Dead roast except where noted otherwise.

[2] First stage is a partial roast in multihearth, second state is a dry-feed dead roast in Dorr-Oliver fluid bed.

[3] Partial roast.

[4] Slurry feed.

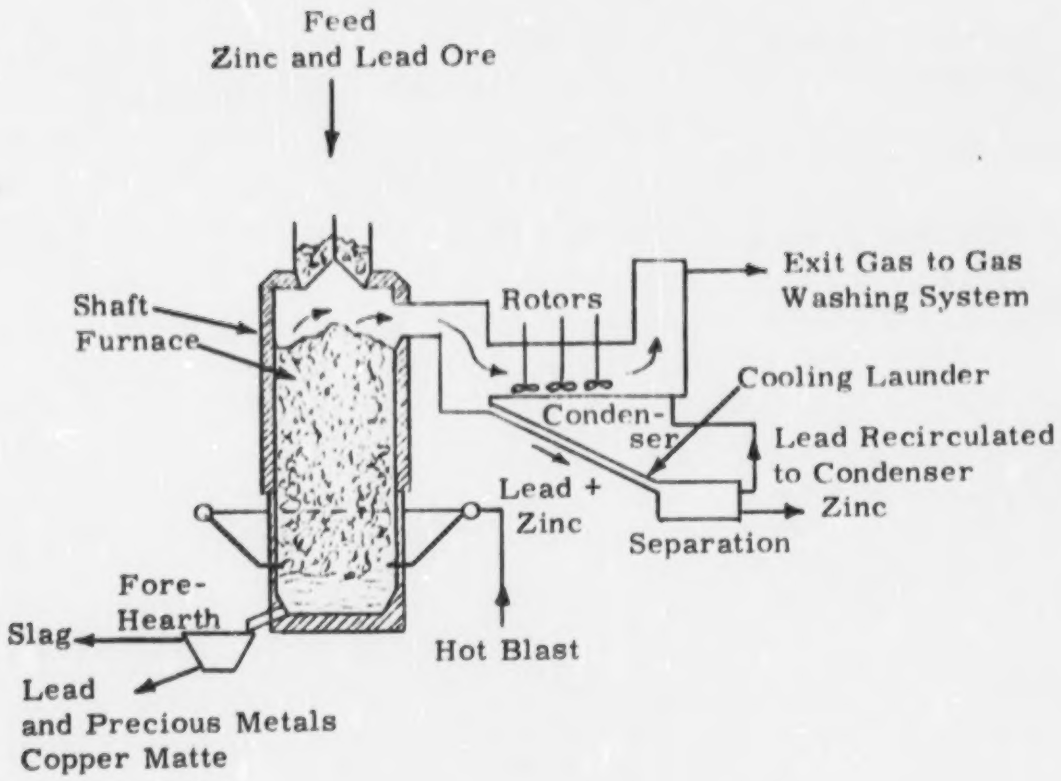
Feed
Zinc and Lead Ore
Shaft Furnace
Rotors
Exit Gas to Gas Washing System
Cooling Launder
Condenser
Lead Recirculated to Condenser
Lead + Zinc
Zinc
Separation
Fore-Hearth
Slag
Hot Blast
Lead and Precious Metals Copper Matte

Figure 22.8. Flow Diagram for a Zinc Plant Utilizing a Retort Process for Zinc Reduction.

Table 22.9

Typical Zinc Sintering Operations

Case:	1	2	3
Total Charge Capacity, tons per day	240-300	400-450	550-600
Machine Size, ft	3.5 x 45	6 x 97	12 x 168
Dust in Off Gas, % of feed	5	5-7	5-10
Off Gas SO_2 Content, %	1.5-2.0	0.1	1.7-2.4

Residues from the zinc processes often contain lead, gold, and silver, which can be economically recovered by subsequent processing.

Dust collection equipment in the production of zinc is usually required on the effluents from dryers, roasters, and sinter machines. The dry processes of dust removal are bag filters and electrostatic precipitators. Wet processes consist of scrubbers. The choice of the type of dust collection equipment depends upon the individual situation and each application has to be decided on its own merit.

22.7 SUMMARY OF PRECIPITATOR OPERATION CONDITIONS

As indicated previously, design of precipitators for the nonferrous metals industry is largely handled by the smelters on a do-it-yourself basis. Table 22.10 lists the basic design data for 3 converter gas precipitators, 3 copper roaster reverberatory furnaces, 1 sinter machine, and 1 acid mist precipitator. Figure 22.9 shows the variation in efficiency with $\frac{A}{V}$ ratio for the 3 converter gas, 2 roaster gas, and 1 sinter machine gas precipitator for which data are available.

22.8 PRECIPITATOR INSTALLATIONS AND ECONOMICS

A summary of available statistics on precipitator performance in nonferrous metallurgical industries is shown in Table 22.11. Statistics on roasters, sinter machines, miscellaneous nonferrous, and the aluminum industry applications are contained in Tables 22.12 through 22.15, respectively.

Available records on these applications are limited, and therefore estimates on total precipitator use would be highly speculative. In general, the aluminum industry offers the most promise for possible increased use of electrostatic precipitators.

Economics. Precipitator cost data for roasters, sinter machines, and the various aluminum applications are presented in Tables 22.16 through 22.18, respectively. Due to the limited amount of data, no attempt has been made to adjust the costs to a common base.

Performance statistics. Due to the limited amount of performance data available and the fact that most of them are old and do not reflect present day state-of-the-art performance, no statistical analysis has been attempted.

Table 22.10

Summary of Electrostatic Precipitators* in Western Smelters Prior to 1920

Precipitator Application	No. of Install.	Volume of Gas Treated (10^3 acfm)	Precipitator Gas Velocity (fps)	Efficiency Range (%)	Comments
Converter Gas	3	519	1.3-5.6	90-99.6	Duct-type-precipitator
Roasters and Reverberatory Furnaces	3	2,475	2.6-15	75-85	Pipe and duct precipitators
Sintering Machine	1	172	8	94.5	Pipe precipitator
Acid Mist (using roaster gases)	1	27	6.5	--	Pipe precipitator
Total	8	3,193			

* Most of these precipitators were built by the smelter owners themselves under patent licensing agreements

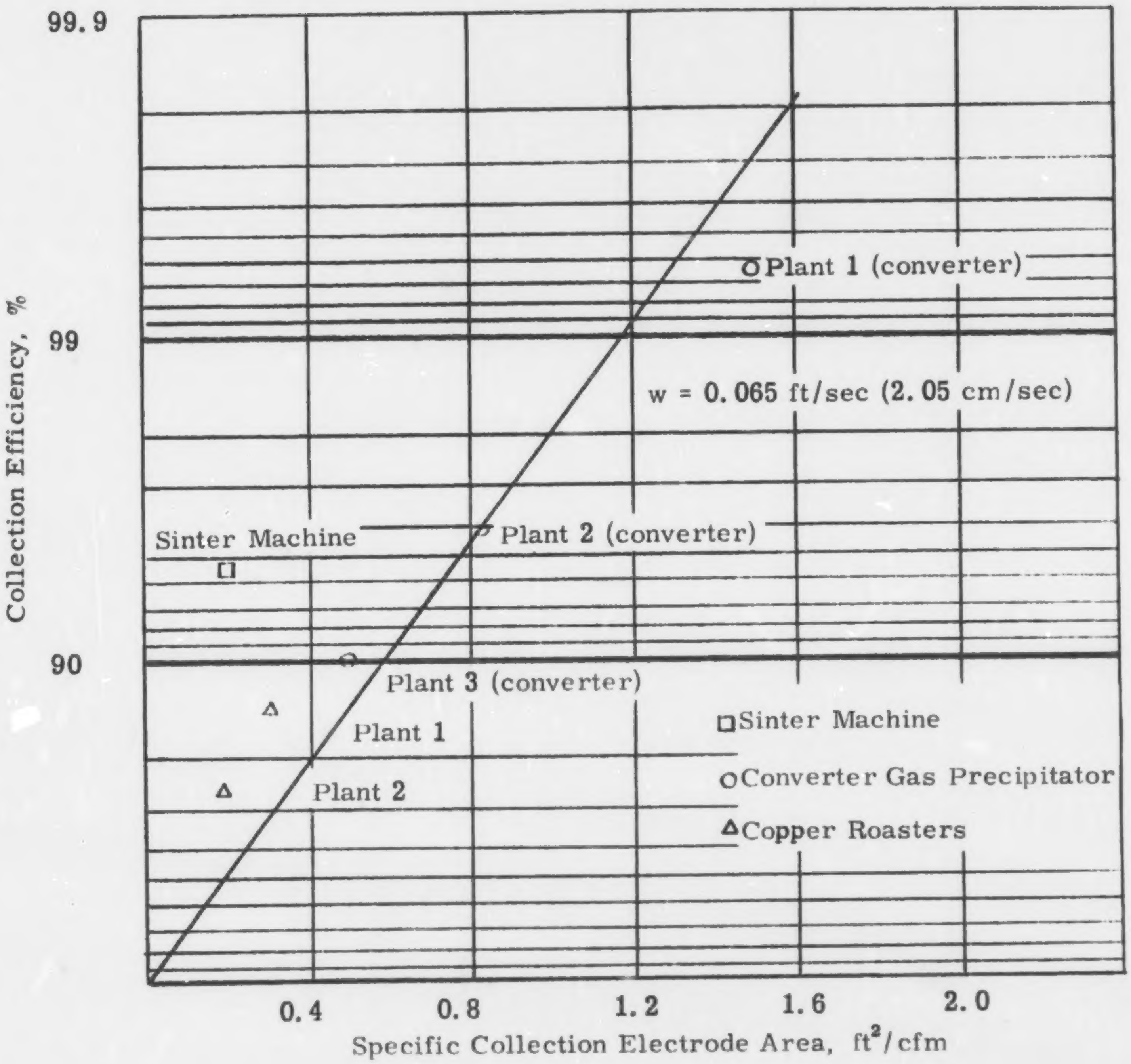

Figure 22.9. Collection Efficiency as a Function of Specific Collection Electrode Area (ft^2/cfm) for Nonferrous Installations.

Table 22.11

Precipitator Performance in the Nonferrous Metallurgical Industries

Precipitator Application	Type	No. of Units or Pipes	Collection Area sq ft	Test Gas Vol 1000 cfm	Test Vol ft/sec	Treatment Time sec	Collection Area sq ft/1000 cfm	Sq ft per Electrical Set	mA per 1000 sq ft	Dust Collection per day tons	Efficiency	Performance w ft/sec
1. Converter Gas												
Plant 1	Duct	20	43,200	28.9	1.3	22	1500	4320		7-12	99.0-99.6	0.05-0.06
Plant 2	Duct	30	95,040	115.0	4.4	12	826	4750			96	0.065
Plant 3	Duct	72	201,960	375	5.6	8	539	4200	11.8	25	90	0.07
2. Copper Roasters and Reverberatory Furnaces												
Plant 1	Pipe	1512 12" x 12'	57,000	350	4.9	2.4	163	6333			75	0.14
Plant 2	Duct	27	110,880	625	11-15	2.2	177	4100	24-94	50-100		
Plant 3	Duct (Vertical Flow)	20	403,200	1500	2.6	8	269	75 kVA sets 40320		5-6	85	0.12
3. Sintering												
Plant 1	Pipe	1800 6" x 12'	33,930	171.8	8.1	1.5	197	15 kVA sets 3393	47	6	94.5	0.24
4. Acid Mist												
Plant 1	Pipe	254 10" x 12'	7,980	27	6.5	3.7	295	3990				

Table 22. 12

Electrostatic Precipitator on Roasters* Period 1923 through 1969

Precipitator Contract Year	No. of Install.	Number of Precipitators	Total Gas Volume (10^3 acfm)	Accumulated Gas Volume (10^3 acfm)	Weighted Design Efficiency on acfm Basis (%)
1923-1929	10	12	260. 7	260. 7	94. 9
1930-1939	15	19	405. 2	665. 9	95. 0
1940-1949	6	9	197. 4	863. 3	96. 8
1950-1959	4	6	200. 6	1063. 9	96. 5
1960-1969	7	12	379. 4	1443. 3	98. 0
Totals	42	58	1, 443. 3		

* This includes hearth, flash, and fluid roasters for zinc, zinc-lead, molybdenum sulfide and pyrites.

Table 22. 13

Summary of Electrostatic Precipitators for Cadmium Recovery Zinc Sinter Machines (1935 through 1955)

Precipitator Contract Year	No. of Install.	Number of Precipitators	Total Gas Volume (10^3 acfm)	Accumulated Gas Volume (10^3 acfm)	Design Efficiency Prorated on acfm Basis (%)
1935-1944	8	19	420. 5	520. 5	95. 0
1945-1955	7	7	594. 5	1015. 0	95. 0
Totals	15	26	1015. 0		

Note: No record of any new installations since 1955.

Table 22.14

Summary of Electrostatic Precipitators for Miscellaneous Nonferrous Applications (*) 1924 through 1968

Precipitator Application	Number of Installations	Number of Precipitators	Total Gas Volume (10^3 cfm)	Design Efficiency Range (%)
1. Tin Smelting	2	5	25	98.0
2. Copper Smelting	3	6	1478	95.0
3. Lead Blast Furnace	5	6	47	90.0-95.0
4. Precious Metals	2	4	208	90.0
Totals	12	21	1758	

* Does not include aluminum industry applications.

Table 22. 15

Summary of Design and Application Parameters on Electrostatic Precipitators for the Aluminum Industry (1940 through 1967)

	DESIGN		RANGE		
Precipitator Application	Temperature of	Gas Volume Per Precipitator (10^3 acfm)	Efficiency (%)	Dust Loading (gr/acfm)	Comments
1. Carbon Plants (Carbon & Coke Dust)	50-175	15-200	98.0-99.0	10-40	Approximately 10 installations
2. Alumina Calciners	350-600	60-200	95.0-99.0	5-10	Approximately 10 installations
3. Aluminum Pot Line	200-225	100-125	90.0-95.0	0.1	Approximately 15 precipitators
4. Remelt Furnace	200-600	10-25	95.0	---	---
5. Soderberg Gases	150-200	50	90.0	0.1	Approximately 30 precipitators

Table 22. 16

Cost Data for Electrostatic Precipitators for Cleaning Ore Roaster Gas

	UNIT COST $/acfm			
Precipitator Contract Year	FOB	Erected	Total Gas Volume (10^3 acfm)	Design Efficiency (%)
1946		1.97	42.0	98
1951		2.17	41.0	98
1953		3.40	37.5	96
1954		6.75	12.0	95
1966	1.90		26.2	99
		6.00	20.0	98
		4.25	47.0	97
1967	0.90		81.5	95
		4.60	33.0	99
1969	1.75		79.2	99

Table 22.17

Economics on Electrostatic Precipitators for Cadmium Recovery Zinc Sinter Machines

	UNIT COST $ /acfm			
Precipitator Contract Year	FOB	Erected	Total Gas Volume (10^3 acfm)	Design Efficiency (%)
1945		0.98	150	95
1946		1.20	100	95
1947		1.18	50	95
1951	0.90		150	95
1953	3.00		12.5	95
1955	2.00		32.0	95

Table 22.18

Economics on Electrostatic Precipitators in the Aluminum Industry

(a) CALCINERS

Precipitator Contract Year	UNIT COST $/acfm FOB	UNIT COST $/acfm Erected	Total Gas Volume (10^3 acfm)	Design Efficiency (%)
1956	0.75		126	98
	0.68		90	95
	0.46		400	98.9
1962		1.10	50	93
1965	0.66		60	78
1967		0.72	70	95
		(b) POTLINE AND SODERBERG GASES		
1951	0.70		1,380	90
1953	1.00		1,320	90
		(c) CARBON PLANT		
1951		1.90	70	98
1952		2.00	10	95

Instead, individual measurements for a number of installations in the roaster and sinter machine applications are presented to indicate the process parameters such as gas volume, gas temperature, dust loading, etc. These data are presented in Tables 22.19 and 22.20.

Design methodology. The procedures used in the selection of types of precipitators for dust control in the noferrrous metals industry are discussed in the sections describing the particular process equipment.

The methodology for determining the collecting plate area is similar to that used in sizing precipitators for other application areas, in that the basic Deutsch-Anderson equation is used. This equation, which relates efficiency to collection plate area, gas volume, and precipitation rate parameter is

$$N = 1 - \exp\left(\frac{A}{v_g} w\right)$$

As in the other application areas, the critical parameter in determining the collecting plate area is the value of the parameter w, which depends upon the resistivity and particle size of the dust, quality of gas flow, reentrainment of the dust, and the particulars of the precipitator design.

As indicated previously, design of precipitators for the nonferrous metals industry is largely handled by the smelters.

Table 22.19

Performance Statistics on Roaster Gas Electrostatic Precipitators (1933-1953)

Installation Number	Gas Volume (10^3 acfm)	Efficiency (%)	Gas Velocity (ft/sec)	Dust Loading (gr/scfd)	Gas Temp (°F)
1	15.5	97.4	2.3	40.0	1100
2	6.25	97.8	1.2	5.1	450
3	15.8	96.5	1.5	8.6	450
4	13.1	95.0	1.4	6.2	540
5	20.0	95.1	4.8	1.9	900
6	18.5	98.8	2.0	15.0	500
7	17.7	98.0	2.9	11.7	900
8	13.3	96.0	1.5	9.5	900
9	18.5	94.9	2.0	7.1	500
10	50.4	86.8	2.8	9.3	500

Table 22.20

Performance Statistics on Cadmium Recovery
Zinc Sinter Machine Gas Precipitators (1935-1955)

Installation Number	Gas Volume (10^3 acfm)	Efficiency (%)	Dust Load (gr/scfd)	Gas Temp (°F)
1	35	97.0	1.59	220
2	35.4	95.8	3.96	205
3	54.3	94.1	1.24	200
4	35.4	97.4	4.99	192
5	26.1	91.7	1.17	250
6	11.0	96.5	0.38	213

CHAPTER 22
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CHAPTER 23
THE APPLICATION OF ELECTROSTATIC PRECIPITATORS TO THE CLEANING OF HIGH PRESSURE, HIGH TEMPERATURE GASES

The cleaning of gases at high temperature and pressure is a unique application for electrostatic precipitators. Above about 600°F, fabric filtration is not suitable because of limitations on the fabric material. Aqueous scrubbers are also not suitable at temperatures much above the boiling point of water. Some consideration has been given to the use of inorganic liquid scrubbers at elevated temperatures; however, costs are generally high so that at the present time there is no commercial use of high temperature inorganic liquid scrubbers.

23.1 HIGH PRESSURE, HIGH TEMPERATURE PROCESSES

There are several processes which require either high pressure or high temperature gas cleaning or a combination of the two. These processes are discussed as special applications in which there may be potential for expanded uses of electrostatic precipitators.

High temperature coal gasification. One of the earliest uses of electrostatic precipitators in high temperature gas cleaning was in the removal of solids from gases produced in the high temperature gasification of coal. Early work in this area was in connection with the development of coal-fired, gas turbine locomotives. The U. S. Bureau of Mines has conducted research and pilot-scale developmental studies of coal gasification over a period of many years.

In the process of gasification, finely divided coal is burned in a fluidized bed chamber that is maintained at high pressure and temperature, typically 1500°F and 600 psi. The gases produced in the process are used to drive gas turbines for electric power generation.

Fly ash generated in the gasification process must be removed before the gas can be used in gas turbines, since the dust would erode the turbine blades.

Pilot-plant gasification systems have been operated by the U. S. Bureau of Mines and gas cleaning by electrostatic precipitators has been demonstrated on these pilot-scale plants. However, at the present time, economic factors do not favor the gasification process, and no full-scale plants have been built.

Waste incineration with gas turbines. A second type of gas cleaning application involving high pressure and temperature is in the cleaning of gas from incinerators in which the gas is to be used to power gas turbines for electric power generation. In many types of incinerators, the disposal of waste material is an exothermic process, and recovery of heat from the effluent gases is economically important. Most heat recovery systems involve the production of steam in waste heat boilers located at the exit of the incinerators. Recent studies indicate that a more economical method of heat recovery may result from the use of effluent gases to drive a gas turbine which, in turn powers an electric power generator. This system requires that the particulate be removed from the gases to prevent erosion of the blades of the gas turbine.

Current studies of the technique of direct turbine drive from incinerator gases are being conducted under contract with the Bureau of Solid Wastes.[1]

The pilot plant system being investigated is shown schematically in Figure 23.1. The refuse is sorted, shredded, and dried in a rotary drier. It is then transported to a pressurized fluid-bed combustion chamber. Combustion air is supplied under pressure from a blower coupled to the turbine shaft. The gas flows upwardly through a combustion zone and then out of the chamber through primary cyclonic type collectors. The fly ash collected in the cyclone is blown out of the reactor and combined with the bottom ash from the fluid-bed chamber.

After leaving the reactor, the gas, under pressure, is passed through an electrostatic precipitator for a final cleaning prior to being delivered to the turbine. Additional heat can be recovered by a waste heat boiler located in the exit gas stream from the turbine.

23.2 INFLUENCE OF TEMPERATURE AND PRESSURE ON PRECIPITATOR PERFORMANCE

Temperature and pressure of the gas have several important effects on precipitator performance. First, gas viscosity increases with temperature. For example, the viscosity of air at 200°F is about 0.02 centipoise and increase to about 0.04 at temperatures around 1200°F. From the equation defining migration velocity, a twofold increase in gas viscosity reduces the migration velocity proportionately. Consequently, one would anticipate reduced performance due to the increase in gas viscosity.

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[1] Refer to the bibliography for this chapter.

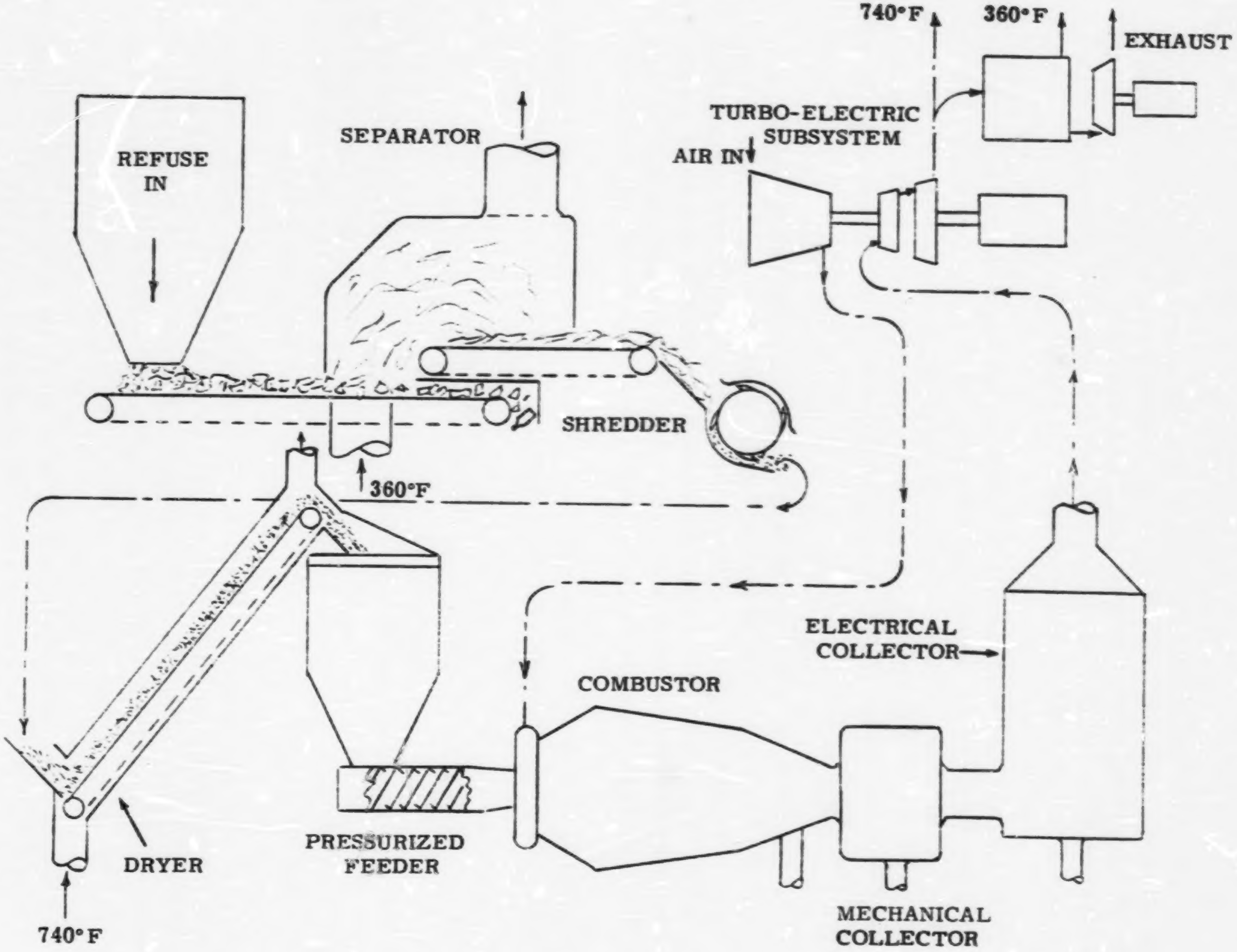
740°F
360°F
EXHAUST
TURBO-ELECTRIC SUBSYSTEM
AIR IN
SEPARATOR
REFUSE IN
SHREDDER
360°F
ELECTRICAL COLLECTOR
COMBUSTOR
DRYER
PRESSURIZED FEEDER
740°F
MECHANICAL COLLECTOR

Figure 23.1. Schematic Diagram of Pilot Plant for Incinerator Gas Turbine System Utilizing Electrostatic Precipitator for Fine Cleaning of Turbine Gas.

A second influence of temperature and pressure on performance is the gas volume. Gas volume is directly proportional to the absolute gas temperature and inversely proportional to pressure. These effects tend to offset each other when high pressure and temperature are both involved, so that gas density is not much affected.

Finally, gas temperature and pressure influence the voltage-current relationships by altering the corona starting voltage and sparking conditions. Increasing temperature tends to lower the corona starting voltage and reduce the voltage at which sparking occurs. Increasing pressure, on the other hand, tends to increase the corona starting potential and the voltage at which the sparking begins.

With negative corona, limitations on current and voltage occur at relatively low temperatures at atmospheric pressure. However, at high pressures, adequate voltage can be maintained to give good precipitation. On the other hand, positive corona generation is not affected to as great an extent and experience indicates that positive corona precipitators may be superior to negative corona at temperatures higher than about 1500°F.[2]

A discussion of the voltage-current characteristics and the influence of temperature and pressure is given in detail in Chapter 1, Part I.

23.3 PRECIPITATORS FOR HIGH TEMPERATURE, HIGH PRESSURE OPERATION

Precipitators for high temperature, high pressure service can be of several types, providing allowance is made for thermal expansion, and for accommodating the high pressures. The latter complicates the problem c dust removal and of providing electrical power to the precipitator.

A pilot unit for high temperature, high pressure dust collection that has been operated in limited tests consists of an 8-inch diameter pipe, 15 feet long with three 0.109 in. diameter wires twisted together as the discharge electrode. The collecting surface area is 31.4 ft^2 with a cross-sectional area of 0.35 ft^2.

Migration velocities of about 0.23 to 0.26 were obtained when precipitating a simulated dust (alumina) with a mass median diameter of about 2.75μ with field strengths (voltage/electrode spacing) of 10-11 kV/in.

Projections to full size units would indicate the following:

Table 23.1

Projections from Pilot Precipitator Tests for Cleaning of Turbine Gas

Gas flow (mass)	210 lb/sec
Temperature	1650° - 1700°F
Pressure	100 psig
Gas velocity	8.25 ft/sec
Gas volume	43,500 acfm
No. of 8 in. precipitator tubes	250
Collecting area	8400 ft^2
Tube Length	16 ft
Length of discharge wire	4000 ft
Voltage	50 kV, avg.
Current	2680 mA
Precipitation rate parameter (w)	0.26 ft/sec

These values are preliminary, based on extrapolation of the pilot-scale unit; however, they indicate the range of values that might be expected.

23.4 CLEANING OF NATURAL GAS IN PIPELINES

An application of precipitators to the cleaning of high pressure gas at low temperatures is in the removal of submicron oil fume from natural gas. In the transmission of natural gas in pipelines, compressors release small amounts of lubricating oil which accumulate in the pipeline and cause loss in transmission efficiency. Manual cleaning of the pipeline is accomplished by passing a plug (called a pig) through the pipeline to remove the oil deposits. The process of pigging is expensive and economics would seem to favor the reduction of the frequency of pigging by removal of the oil mist. In the past, filters, scrubbers, and cyclones have been used in this application. However, high pressure electrostatic precipitation has been used in this application, with apparent satisfaction.[3]

Table 23.2 shows the precipitator design and operating conditions for gas pipeline cleaning.

Table 23.2

Gas Pipeline Precipitator
Design and Operating Data

Gas composition	94% methane
Gas temperature	90° - 130°F
Gas pressure	830 - 860 psi
Gas viscosity	1.2×10^{-4} poise
Gas density ratio (basis 60°F, 15 psi)	51.5 av.
Gas density (specific gravity = 0.6 relative to air)	2.36 lb/cu ft
Gas flow rate, millions of scf per day	300 - 400
Gas flow rate at standard conditions	208,000 - 278,000 scfm
Gas flow rate at conditions	4,000 - 5,400 cfm
Collection efficiency	99% by weight
Oil mist collected	9 - 10 gal/d max
Particle size	0.1 - 1 micron
Installation	2 parallel pptrs on 26" pipeline
Pressure vessels	ASME code, 900 psi; hydraulic test, 1800 psi
Particle concentration	7×10^{-5} to 2.4×10^{-4} gr scf (0.0034 - 0.012 gr acf)
Precipitator voltage	55 - 58 kV av.
Precipitator current	80 - 90 mA DC
Corona power	1 watt/cfm
Duct Reynolds No.	1.4×10^{6}
Pressure drop	0.1 psi
Power input	12 kVA

One installation of a precipitator for natural gas pipeline was made in 1967. It is a horizontal-flow, duct-type consisting of two parallel units, each unit consisting of two 5-foot sections 2 feet high, with 3-inch duct spacing. Each precipitator unit is contained in a 36-in. diameter pressure vessel shell designed for 900 psi at 140°F. Conical inlet and outlet transition sections approximately 6 ft long connect to the 18-in. diameter pipe.

The precipitator is equipped with a wash-in-place system composed of a pipe manifold and spray nozzles above the plates. The spray system capacity is 48 gal/min at 10 psi. Kerosene-type hydrocarbon fluid is used for washing.

The discharge electrode system consists of wires, 10 mils in diameter, strung under tension between a top and bottom support frame. The wires are centered between adjacent collecting plates. The high voltage discharge frame is supported by insulators located in and fastened to the sides of the precipitator.

The collecting electrodes are flat metallic sheets carried by top and bottom spacer bars. Openings are provided in the plates to accept the passage of the high voltage support members. The downstream end of each plate is equipped with a shield in order to eliminate any reentrainment of oil from the end of the plate.

Energization system for pipeline precipitator. The high voltage energization system consists of a dry-type, encapsulated, high-voltage, step-up transformer, coupled with a full-wave, solid-state rectifier, and is provided with a fully automatic control system. The power pack is contained in a compartment appended to the precipitator and is exposed to full gas line pressure. The low voltage line controls are located externally to the pipe lines.

Since the precipitate is free flowing, removal consists of draining the precipitator pan to an appended pressurized sump from where the oil is periodically blown out by pipeline pressure.

Economics. The FOB cost of pipeline precipitators to handle 2×10^9 scf per day of natural gas at 800 psia at 98.5% collection efficiency is reported to be on the order of $10/acfm.

CHAPTER 23
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CHAPTER 24
NEW PRECIPITATOR APPLICATIONS

The choice of the type of dust control equipment for any gas cleaning application depends upon economic and process factors. In many instances, a particular type of dust control equipment may historically have been utilized for certain types of service, and trends are often followed when control equipment is selected for an application in which there has been a history of successful performance in similar service. In a number of application areas, a variety of types of dust control equipment has been used successfully, and the choice in such cases is dependent upon local factors or often upon personal preferences.

The major types of equipment for control of industrial dust emissions are mechanical collectors, fabric filters, electrostatic precipitators, and wet scrubbers. Mechanical or inertial collectors can be of the low-loss type such as cinder traps, which are used to prevent fallout of large particles in the vicinity of the plant, or of the high-draft-loss type for higher collection efficiency.

Mechanical cyclone collectors are of the high-draft-loss type and are used on particles in the 10-200μ range. The efficiency of cyclone collectors varies with the size of the dust, and decreases rapidly for particles under 10μ diameter. Figure 24.1 shows the collection efficiency of a mechanical cyclone collector as a function of particle size and pressure drop.[1] Mechanical collectors are often used in combination with electrostatic precipitators for control of fly ash from power plant boilers.

Wet scrubbers remove particulate from a gas stream by impinging the gas against water fog or a wetted surface. The simplest type of scrubber involves a water spray to provide small diameter liquid particles which contact the dust-laden gas. On contact, the dust particles are transferred from the gas stream to the scrubbing liquid and are removed from the gas along with the liquid under the influence of centrifugal or gravitational forces. Simple spray chamber or packed tower scrubbers are inefficient for removal of small size dust.

For small particles, high energy wet scrubbers are used to achieve high collection efficiencies. The collection efficiency for a given size is

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[1] Refer to the bibliography for this chapter.

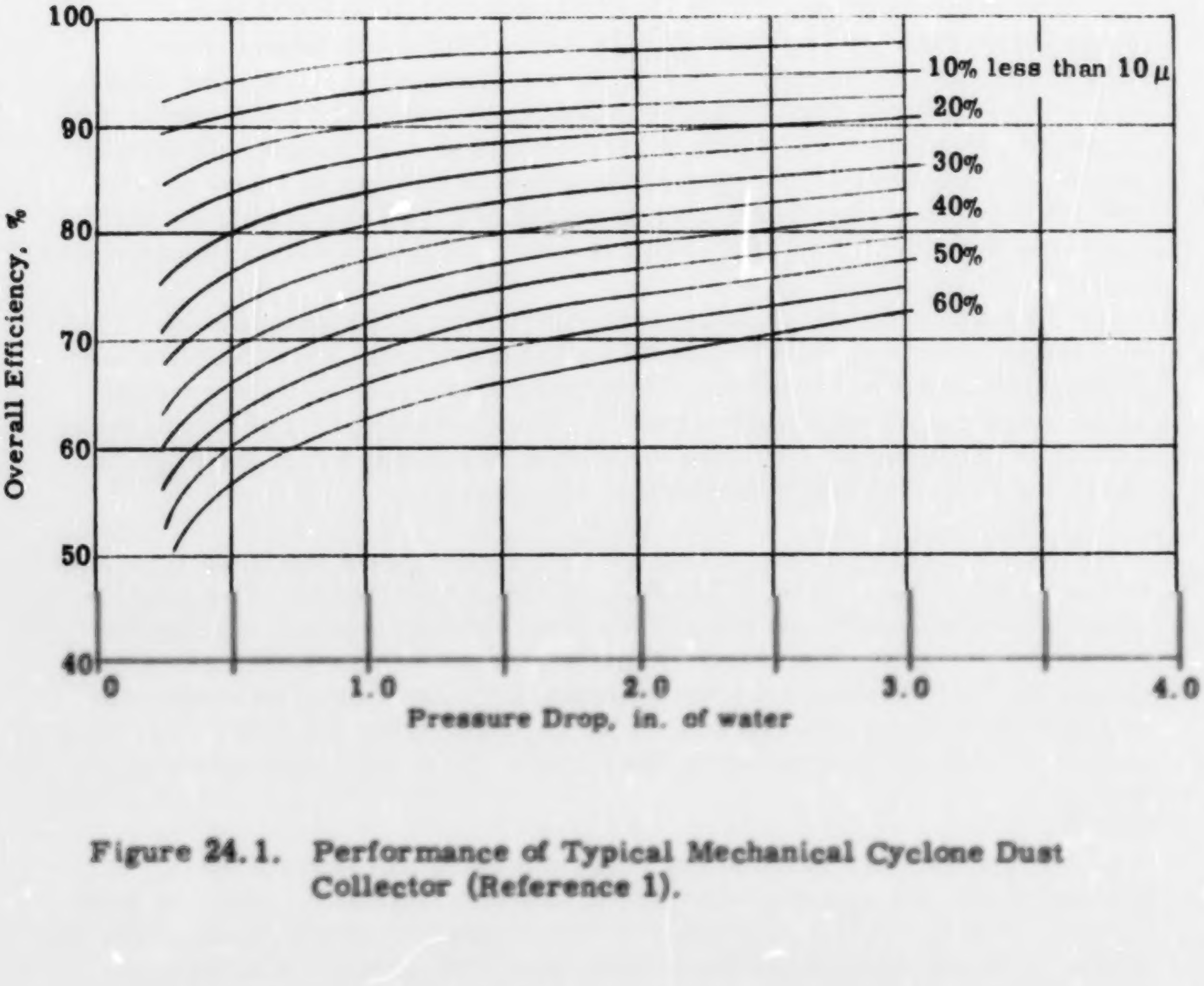


Figure 24.1. Performance of Typical Mechanical Cyclone Dust Collector (Reference 1).

directly related to the gas pressure drop across the scrubber, and for high collection efficiencies of small particles, pressure drops of 3-100 in. water are used. An added advantage of the wet scrubber is that gas absorption can also occur during the process of particulate removal. Major difficulties involved in wet scrubbing lie in high operating costs if large gas volumes are to be handled, disposal of the scrubbing liquid or sludge, and loss of plume buoyancy. Steam plumes formed from scrubbing hot gases are a further disadvantage of wet scrubbers, and if steam plume suppression is required, costs are often high.

The power required to move gas against a given pressure drop is given by

$$P = \frac{v \times p \times 746}{6356 \times \eta}$$

where

v = gas volume handled, cfm,
p = pressure drop, in. water,
P = input power, kW, and
η = *combined fan and motor efficiency.*

For a gas volume of 1×10^6 cfm and a 10 in. pressure drop, the power required for a 50% efficient fan and motor combination would be

$$P = \frac{1 \times 10^6 \times 10 \times 746}{6356 \times 0.50} = 2360 \text{ kW}.$$

Fabric filters are intrinsically high efficiency collection devices, and are particularly suited to controlling dusts where low gas volumes are handled. Fabric filters are limited to temperatures less than about 600° F, to effluents that do not chemically attack the fabric, and to dusts that do not clog or foul the collectors.

Economic considerations in the selection of dust control equipment include capital investment, operating costs, and maintenance costs of the dust control facility.

The installed or capital cost comparisons for a particulate control system must be carefully made for the specific application. They include the cost for the control device itself, installation costs, and cost of any auxiliary equipment that may be required. Edmisten, et al.,[2] presents typical capital cost data for low, medium, and high collection efficiencies of dry centrifugal collectors, wet collectors, electrostatic precipitators, and fabric filters. These are given in Figures **24.2** through **24.5**. Figure **24.6** shows installed cost of electrostatic precipitators as a function of gas volume handled.

Installation, operating, and maintenance costs for the various types of collectors are summarized in Table **24.1**.

Figure **24.7** is a comparison of the total accumulated costs of an electrostatic precipitator, fabric filter, and wet scrubber, for collection of a particular fine dust. The procedure used in calculating the costs is as follows:

Total Accumulated Cost = FOB Cost + Installation Cost + (Annual Operating Cost + Annual Maintenance Cost) (t Years)

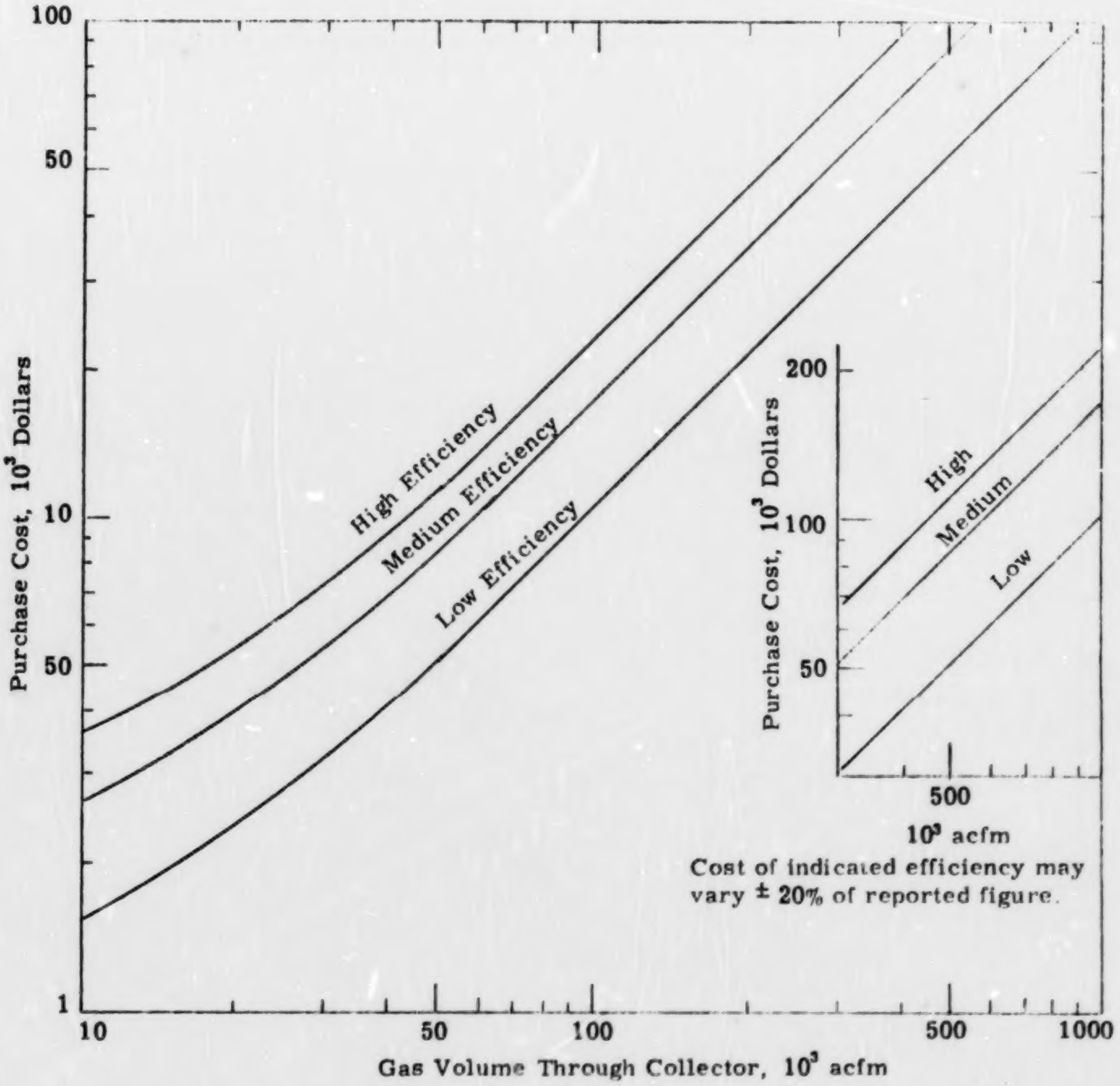
Cost (Electrostatic Precipitation) = **$136,000 + $ 5,500 t**

Cost (Fabric Filter) = **$110,500 + $24,500 t**

Cost (Wet Scrubber) = **$ 37,500 + $51,500 t**

Data used in the calculations are those of Walker[3] and are shown in a comparative breakdown in Table **24.1**. The total costs are for a small **(100,000** cfm) installation.

The economic considerations in the selection of dust control equipment should be resolved for a particular application based on analyses which indicate the long-term costs of comparative dust control equipment. The cost analyses should be carefully studied to insure that they are made on a comparable basis. Generalized cost data, even within a given application area, can be in error by a considerable margin if dust and gas properties vary, as they often do. Consequently, the cost analyses



Cost of indicated efficiency may vary ± 20% of reported figure.

Figure 24. 2. Purchase Cost of Dry Centrifugal Collectors (Reference 2).

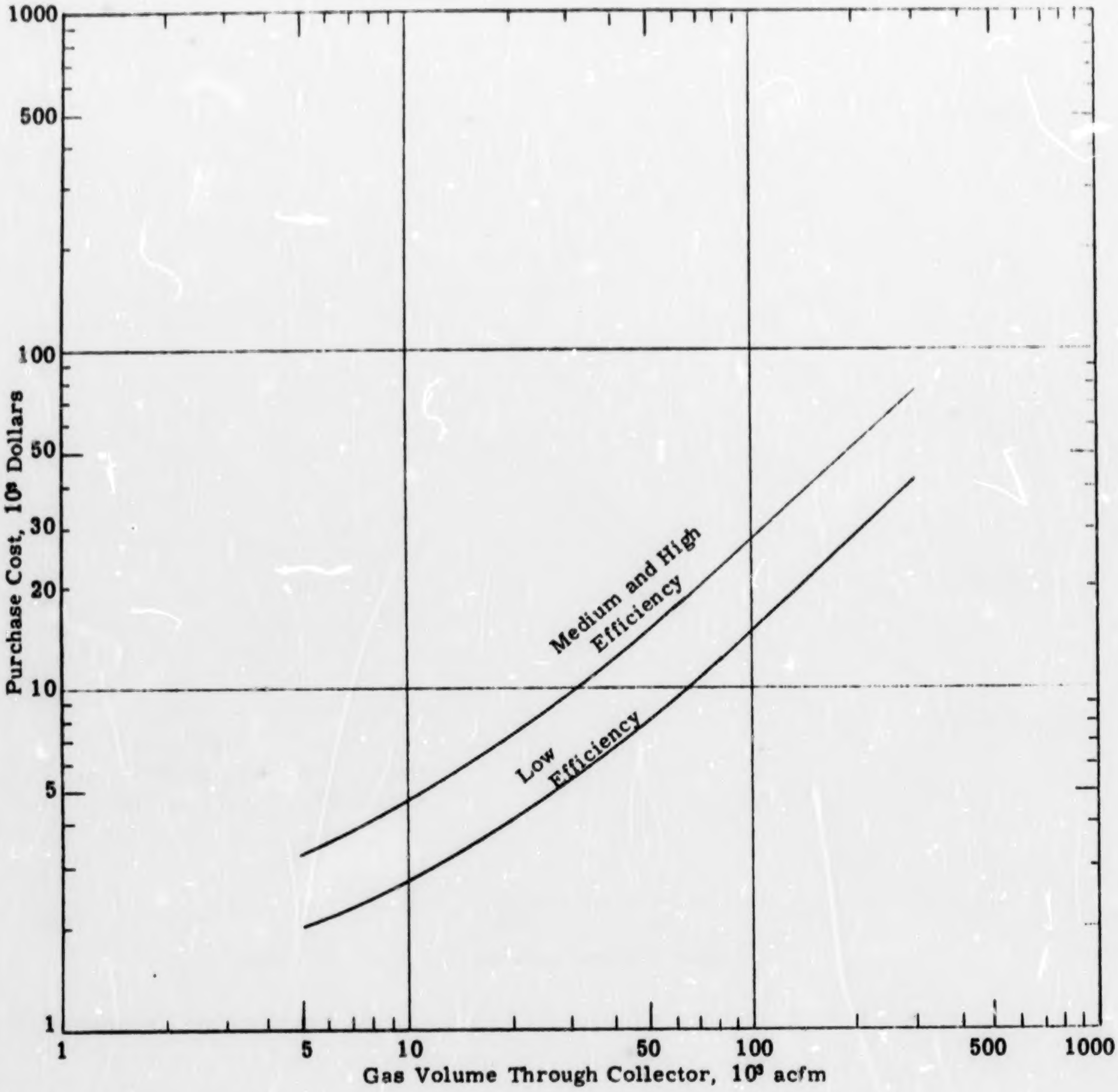


Figure 24.3. Purchase Cost of Wet Collectors (Reference 2).

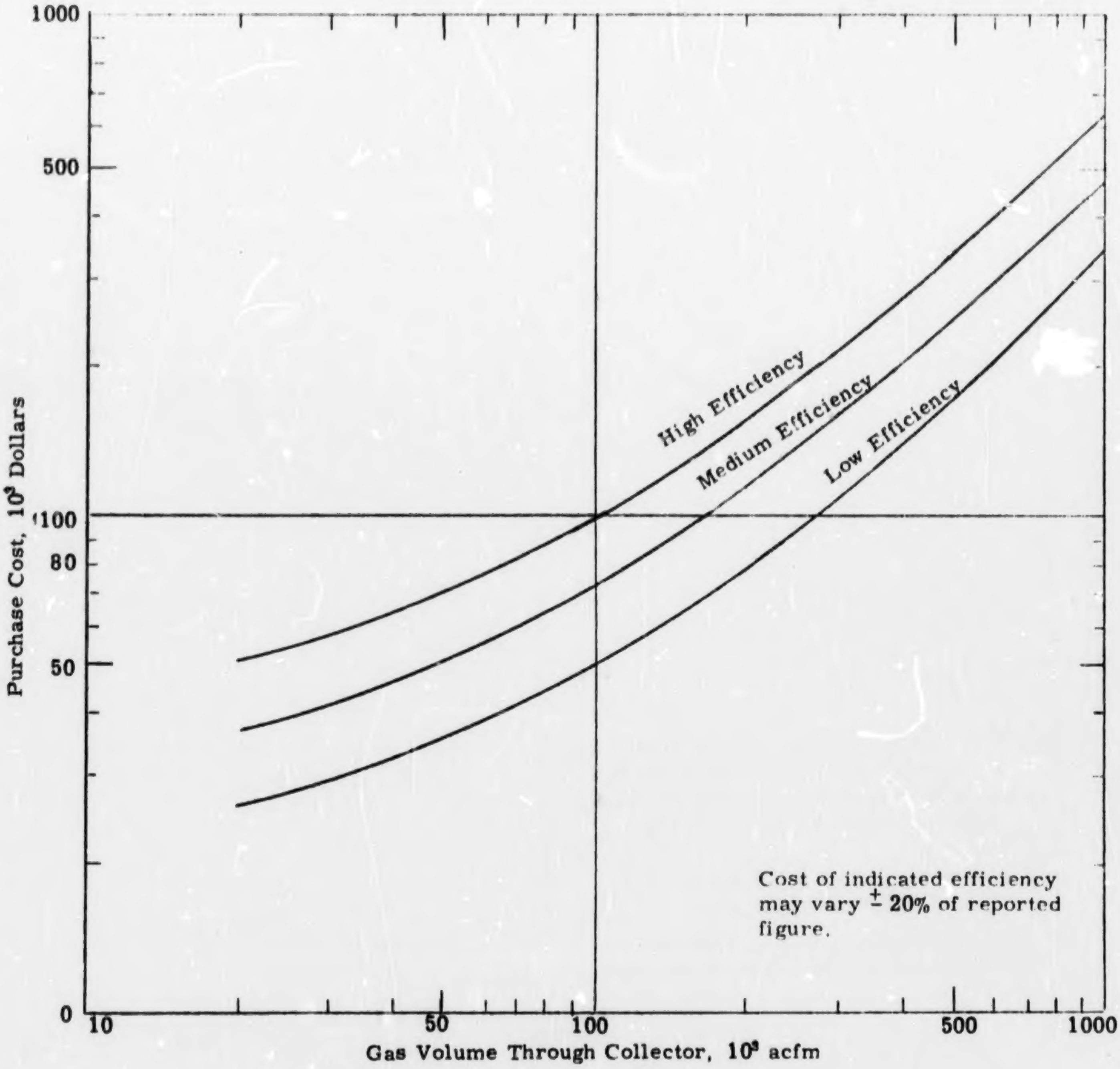


Figure **24.4.** Purchase Cost of High-Voltage Electrostatic Precipitators (Reference 2).

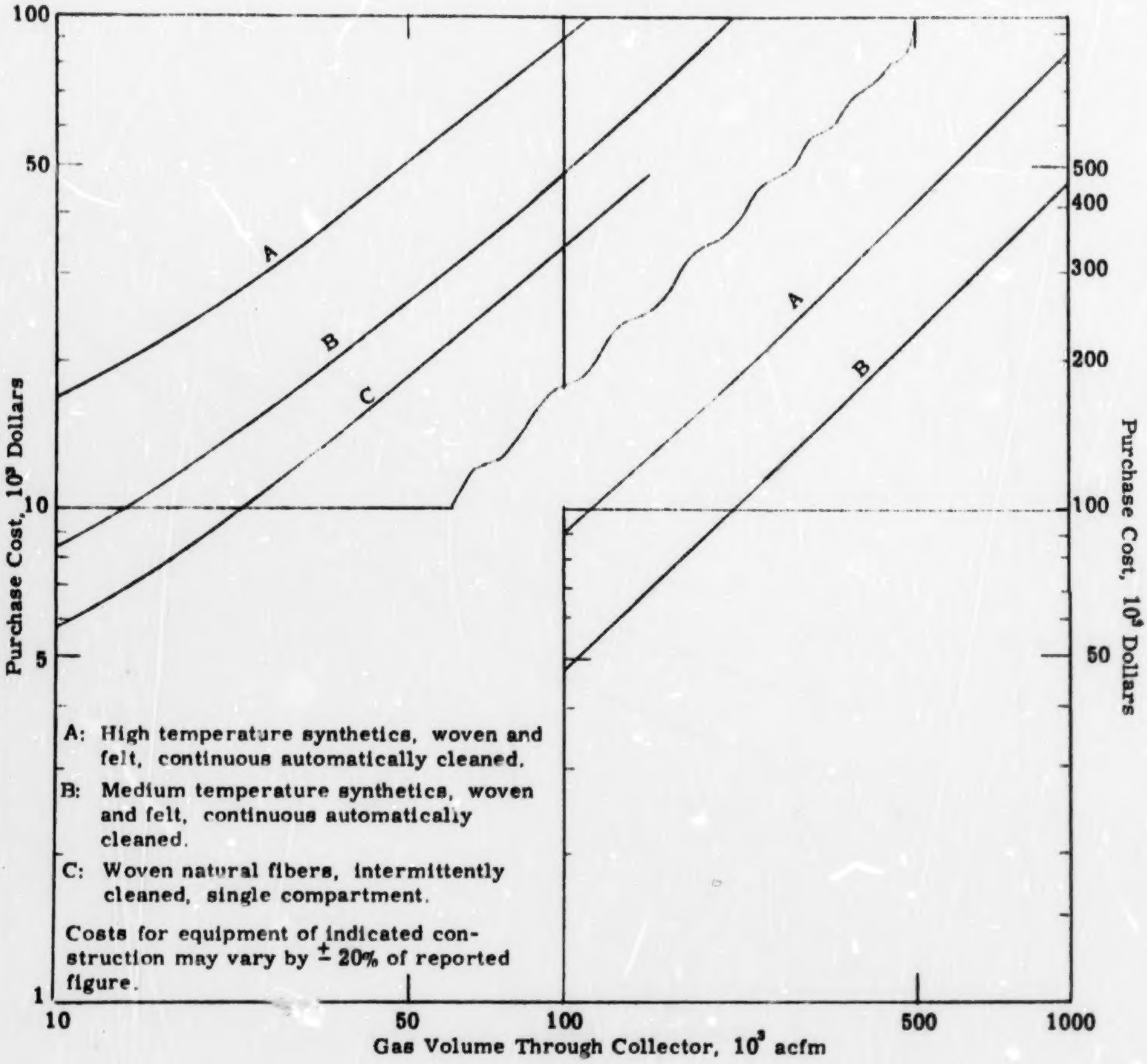


Figure 24. 5. Purchase Cost of Fabric Filters (Reference 2).

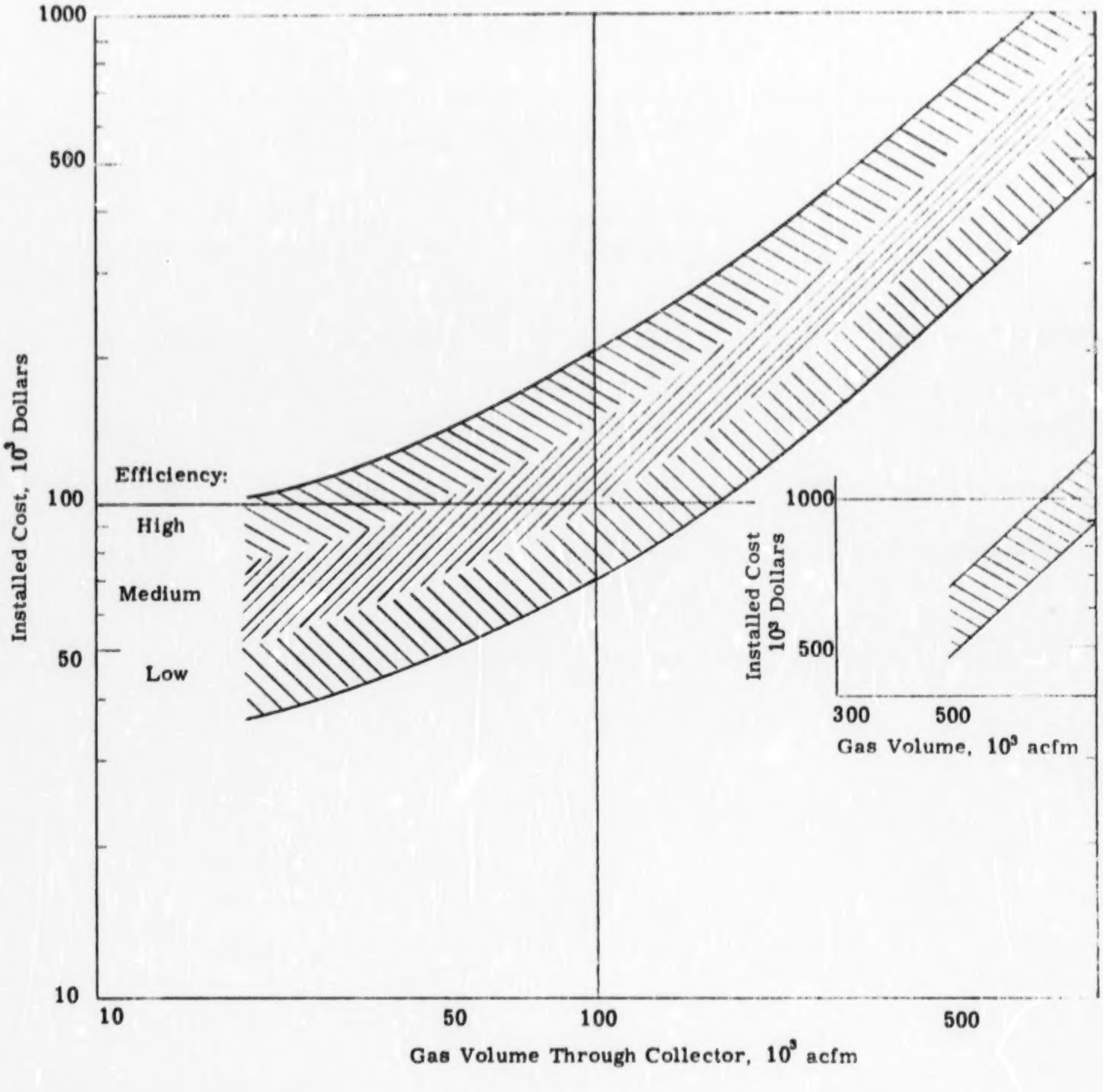


Figure 24.6. Installed Cost of High-Voltage Electrostatic Precipitators (Reference 2).

Table 24.1

Comparison between Particulate Control Costs for a 100,000 cfm Unit (Ref. 3)

	Electrostatic Precipitator	Fabric Filter	Wet Scrubber*
FOB Cost	$80,000	$85,000	$30,000 (Mild Steel)
Installation	56,000	25,500	7,500
Operating Per Year	3,000	7,000	50,000
Maintenance Per Year	2,500	17,500	1,500

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* Does not include clarifiers and liquid treatment systems.

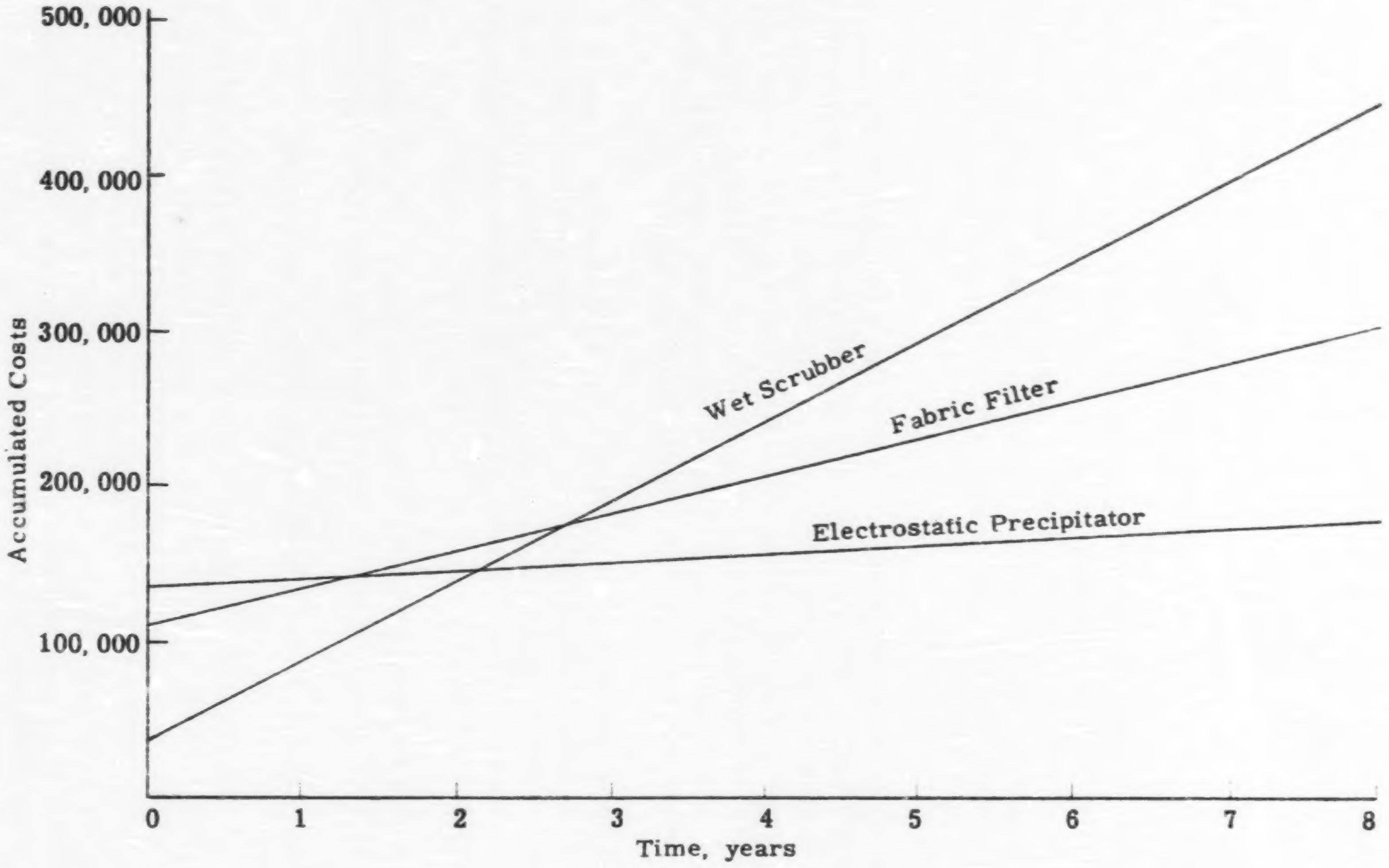

Figure 24. 7. Comparative Costs for Particulate Control for 100, 000 cfm Installation

should be based on specific data for a given plant as opposed to a generalized analyses. For example, the initial cost of an electrostatic precipitator is directly related to the collection surface area required. This surface area requirement is, in turn, a function of the precipitation rate parameter, which can vary by factors of 2 or more depending on dust resistivity and particle size. Hence, the initial cost of an electrostatic precipitator can be in error by a like amount.

As another example, the energy requirement to achieve a given efficiency in a wet scrubber is dependent on particle size. Thus, rather severe changes in operating costs could result from an error in particle size assumptions. As a generality, the accuracy of a particular cost estimate is directly related to the time, effort, and ability put into its preparation.

Factors other than cost also influence the selection of the type of dust control equipment. An electrostatic precipitator normally operates in a sparking condition, so that combustible materials are not generally candidates for control by precipitators. Applications of this type include feed or flour mill effluents and coal grinding dusts. However, precipitators are used in collection of tar and other materials where air can be excluded to prevent the buildup of a combustible mixture.

Fabric filters are normally limited to applications with gas temperatures less than about 600° F due to heat damage to the types of fabrics available. Some types of dust are also difficult to collect with fabric filters because of their tendency to clog the fabric and increase the pressure drop excessively.

Wet scrubber applications are limited to those in which disposal of the slurries can be achieved without creating a serious pollution problem.

24.1 APPLICATION OF ELECTROSTATIC PRECIPITATORS

The use of electrostatic precipitators is generally best suited to those areas in which large gas volumes are to be handled, and where high efficiency collection of relatively small particles is desired. The use of electrostatic precipitators for particulate emission control has steadily increased since it was first introduced for industrial dust control. Figure 24.8 shows the total shipments of various types of particulate control devices for 1963 and 1967. As can be seen, the 1967 shipments of electrostatic precipitators were about 2.6 times those of 1963, indicating a substantial growth pattern.

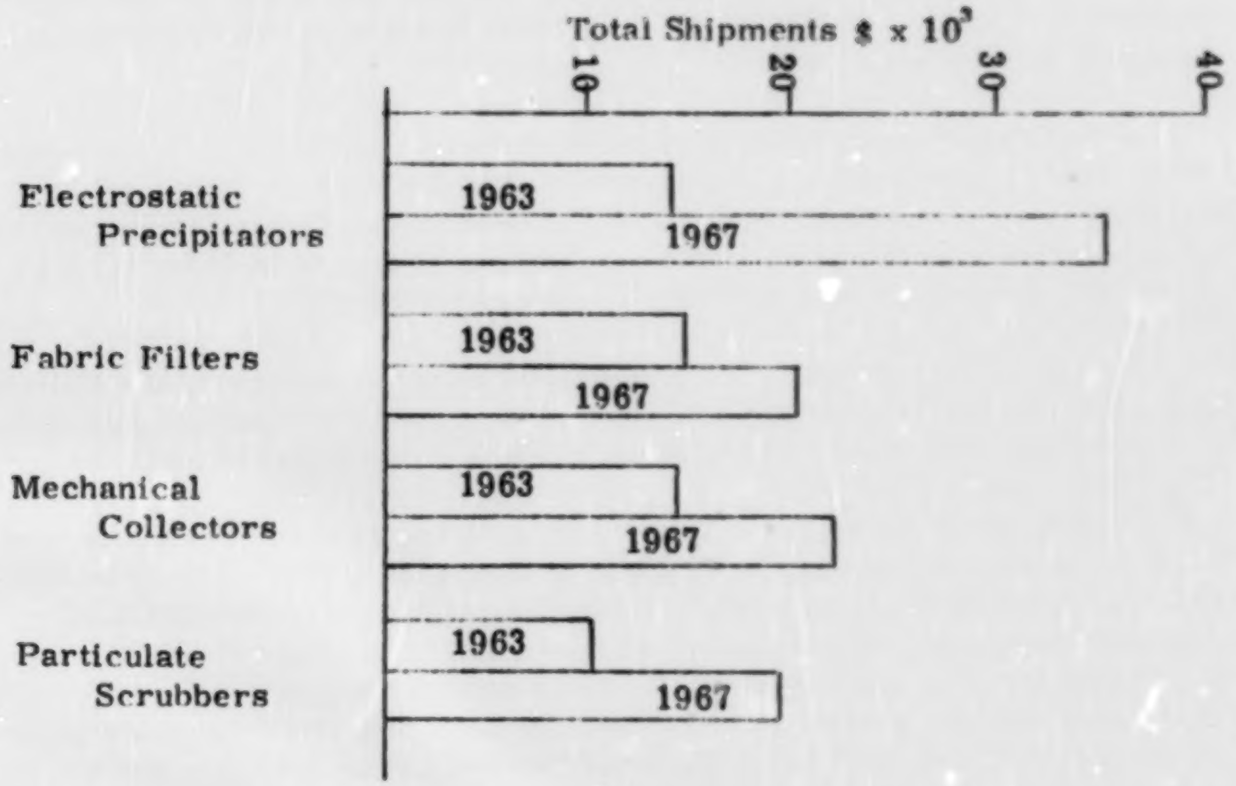

Figure 24.8. Total Shipments of Various Types of Particulate Control Devices for 1963 and 1967.

This growth, however, is due principally to the large usage in coal-fired electric power generating plants. The rapid expansion of the power generating capacity in this country has been accompanied by corresponding increases in precipitator sales to that industry since precipitators, either separately or in combination with mechanical collectors, are the only type of dust control equipment used to any significant degree.

Table 24.2 is a compilation of the number of electrostatic precipitators installed in all of the applications in which they are used in this country. This table indicates several trends in the application of electrostatic precipitators in which changes in the process itself or the introduction of new control systems have altered the use of electrostatic precipitators to a substantial degree.

The use of electrostatic precipitators to control emissions in the major application areas is discussed in the chapters covering those applications. Trends in the type of dust control equipment used, and general trends in the growth of the industry, are also included as they affect the future use of electrostatic precipitators.

In terms of new uses for electrostatic precipitators, possibilities exist for the application of precipitators where other control techniques are used and in areas where no dust control is currently practiced.

Table 24.3 shows a breakdown of sales of dust control equipment used in various processes during the period 1966-1967.[4] The processes are identified by SIC categories. The figure clearly shows the major areas in which the various types of dust control equipment are used and the relative use of each type of control equipment in each process. Opportunities for the use of electrostatic precipitators in areas where other types of dust control equipment predominates depend primarily upon economic factors. There are, however, as discussed previously, other considerations on which choice is made.

One of the major limitations of electrostatic precipitators is the difficulty in handling high resistivity dusts. If the dust resistivity is higher than about 5×10^{10} ohm-cm, a condition of reverse ionization or back corona occurs which seriously impairs the collection efficiency. To minimize the effect, the collecting plate area must be substantially increased to reduce the current density, or other corrective measures must be taken which can result in high costs and make electrostatic precipitators less competitive in certain applications.

Table 24.2

A Summary of Electrostatic Precipitator Application

Application Area	1920-1924	1925-1929	1930-1934	1935-1939	1940-1944	1945-1949	1950-1954	1955-1959	1960-1964	1965-1969	Total
Fly Ash	1	11	12	56	74	90	63	90	87	171	655
Metallurgical											
Ore roasters, zinc or lead zinc		9^1	2	8	8	2	2	1		1	33
Ore roasters, molybdenum sulfide		2^1	1	1	1						5
Ore roasters, pyrites		2	1	3	1		1				8
Aluminum (carbon plant)					2		3				5
Soderberg pots					2			1			3
Aluminum remelt furnaces							3	2			5
Aluminum prebake potline							3	2			5
Cryolite recovery							1				1
Cadmium recovery zinc sinter mach				5	5	4	4				18
Copper				1						1	2
Lead blast furnace		3^1	3		1						7
Gold and silver			1	2			1			2	6
Tin smelting		1			2						3
Scarfing machine								4	3	1	8
Sintering machine							1	7	7^2		15
Blast furnace			4	17	29	24	26	21	6		127
BOF									2	5	7
Open hearth							2	9	8	4	23
Sinter machine gas							2	2		3	7
Chemical											
Sulfuric acid							15	10	1	4	30
Acid mist							3	12	5	8	28
Phosphorus electric furnace					2	2	2	1	2	1	10
Phosphate							2				2
Phosphoric acid mist		2		1	3	1	2	1			10
Disodium metaphosphate							1				1
Rock Products											
Alumina calciner					3		3			1	7
Cement kiln (wet)				7^3		1	7	24	17	17	73
Cement kiln (dry)				14^3			3	17	6	3	43
Cement plant dryers and mills				1^3				2	3		6
Gypsum				4	3	3	18	19	10	1	58
Petroleum											
Catalytic crackers					24	13	3	2			42
Pulp and Paper											
Pulp mill					12	28	20	37	2	15	114
Paper mill						2	2	11	25	31	71
Tar Recovery		63^1	20	33	62	82	35	14	6		282

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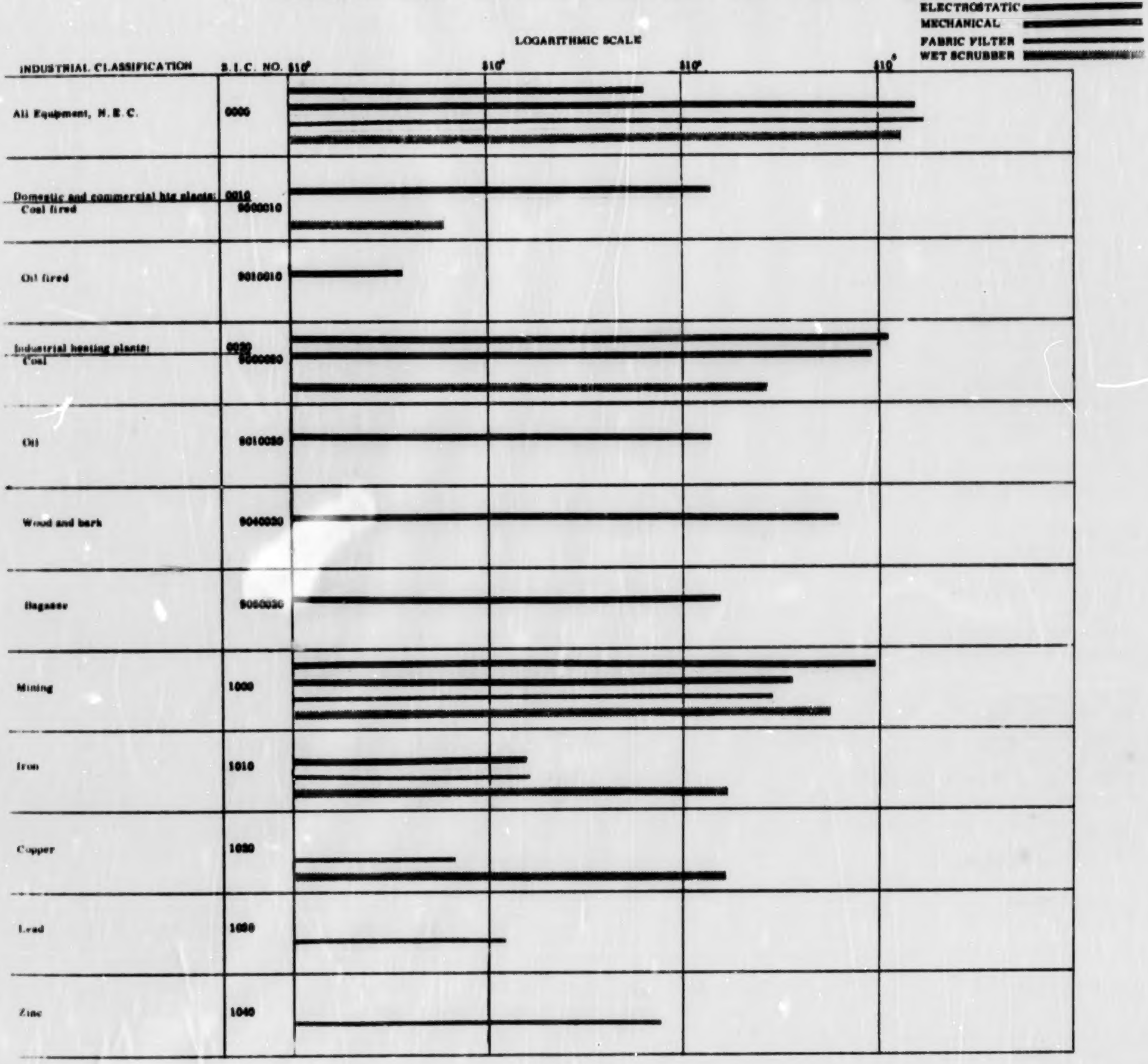
[1] Prior to 1930
[2] 1960-1969
[3] 1930-1939

TABLE 24.3

SALES OF DUST CONTROL EQUIPMENT USED IN VARIOUS PROCESSES DURING THE PERIOD 1966-1967



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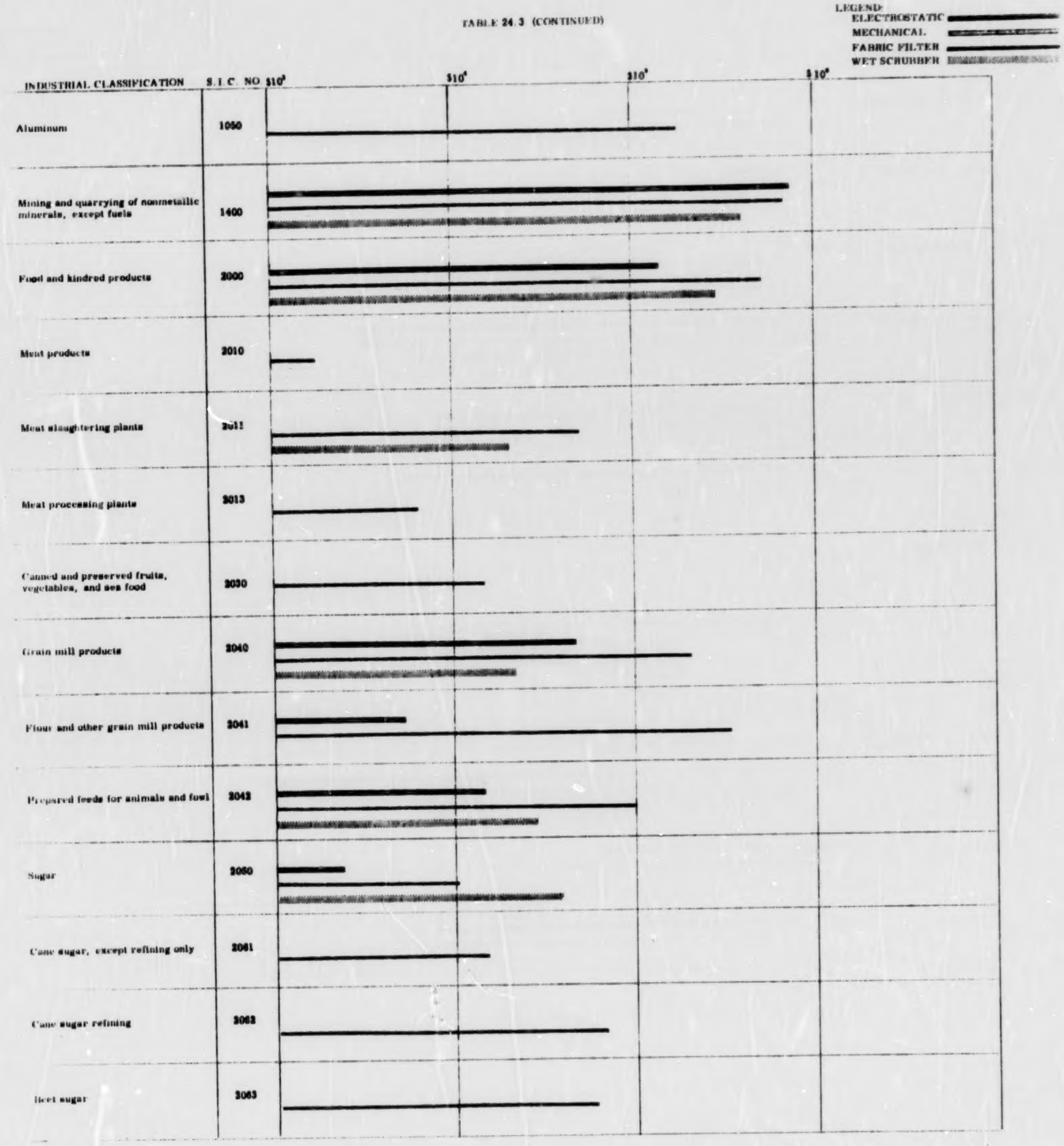

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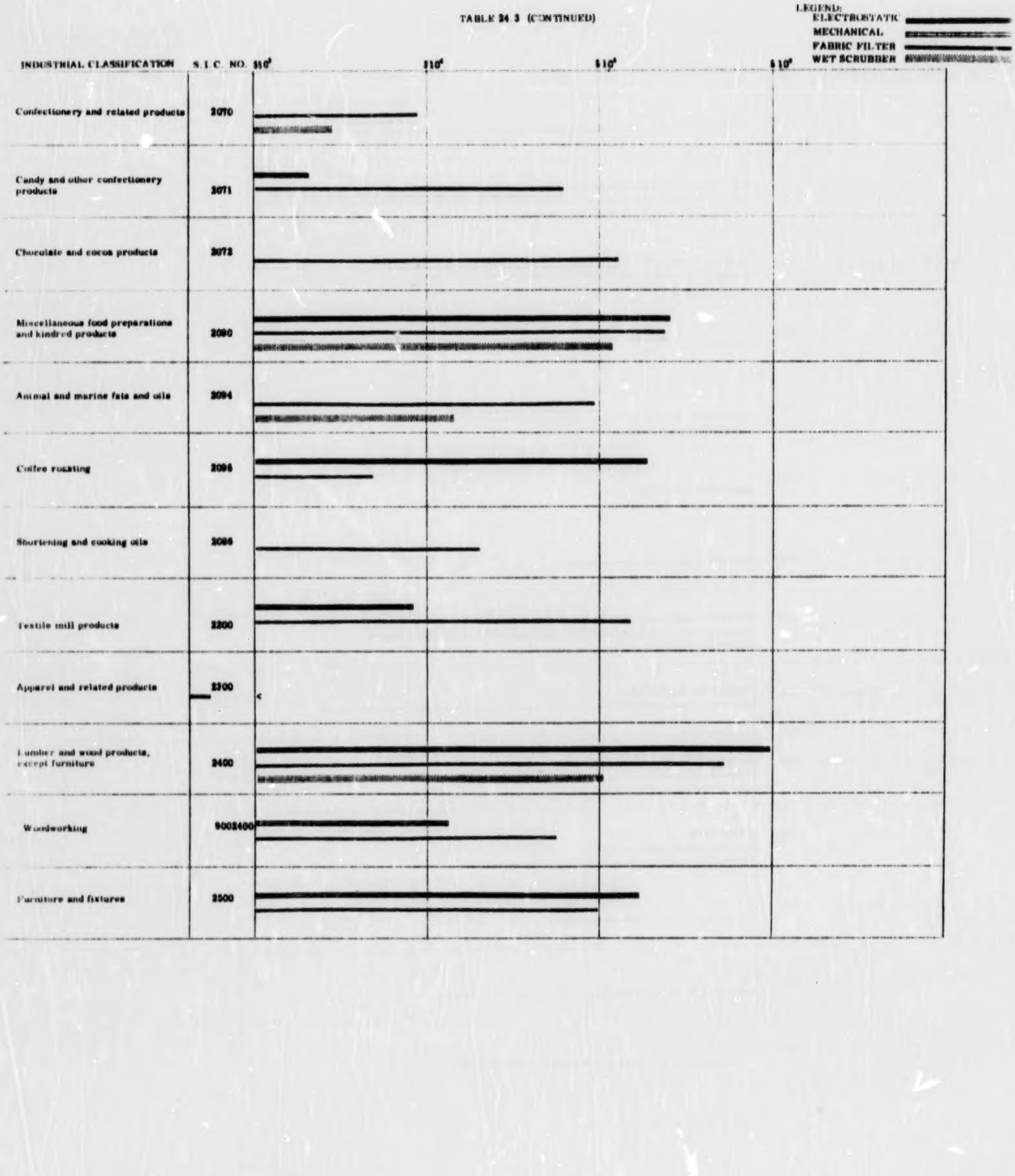

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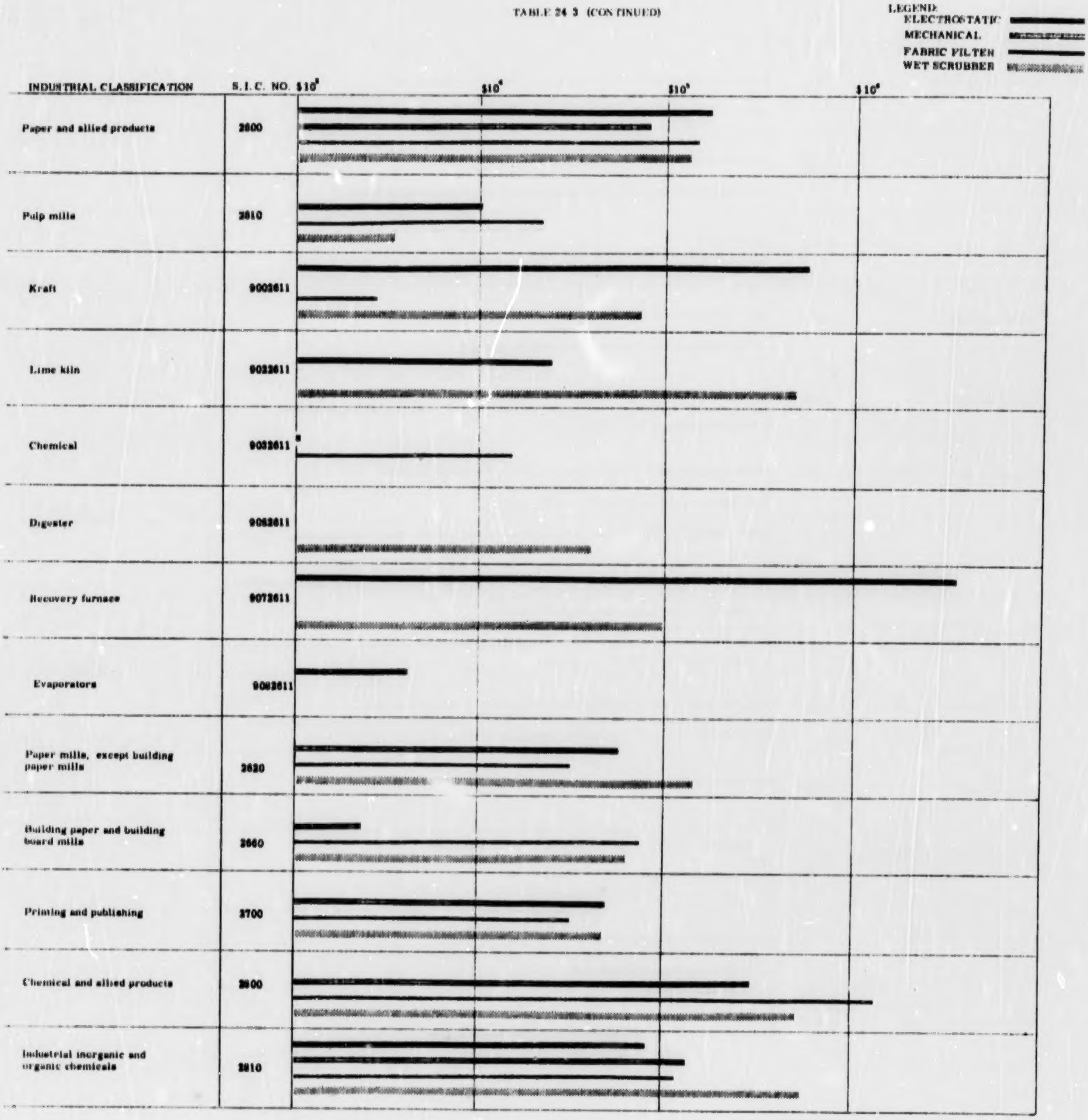

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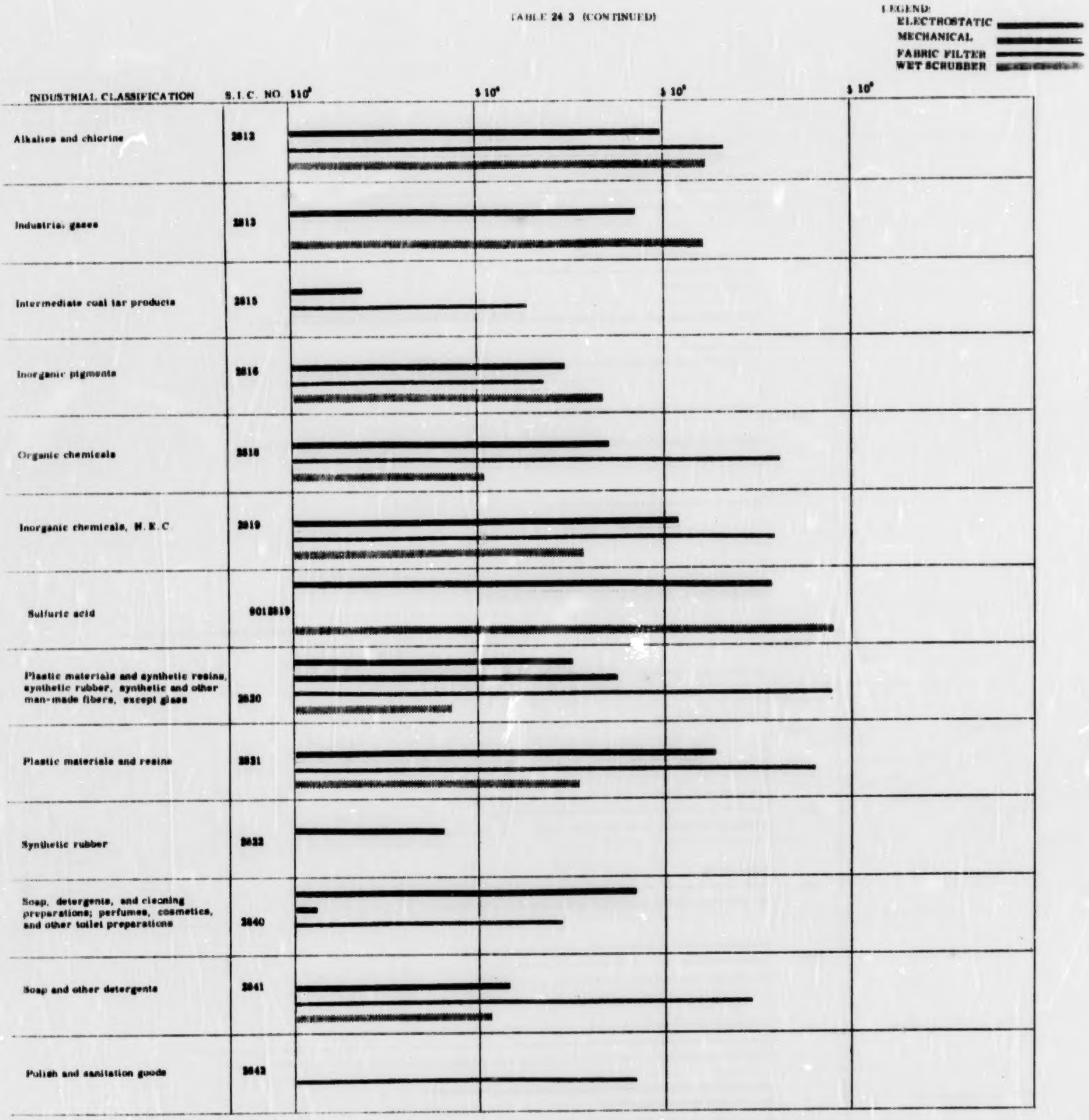

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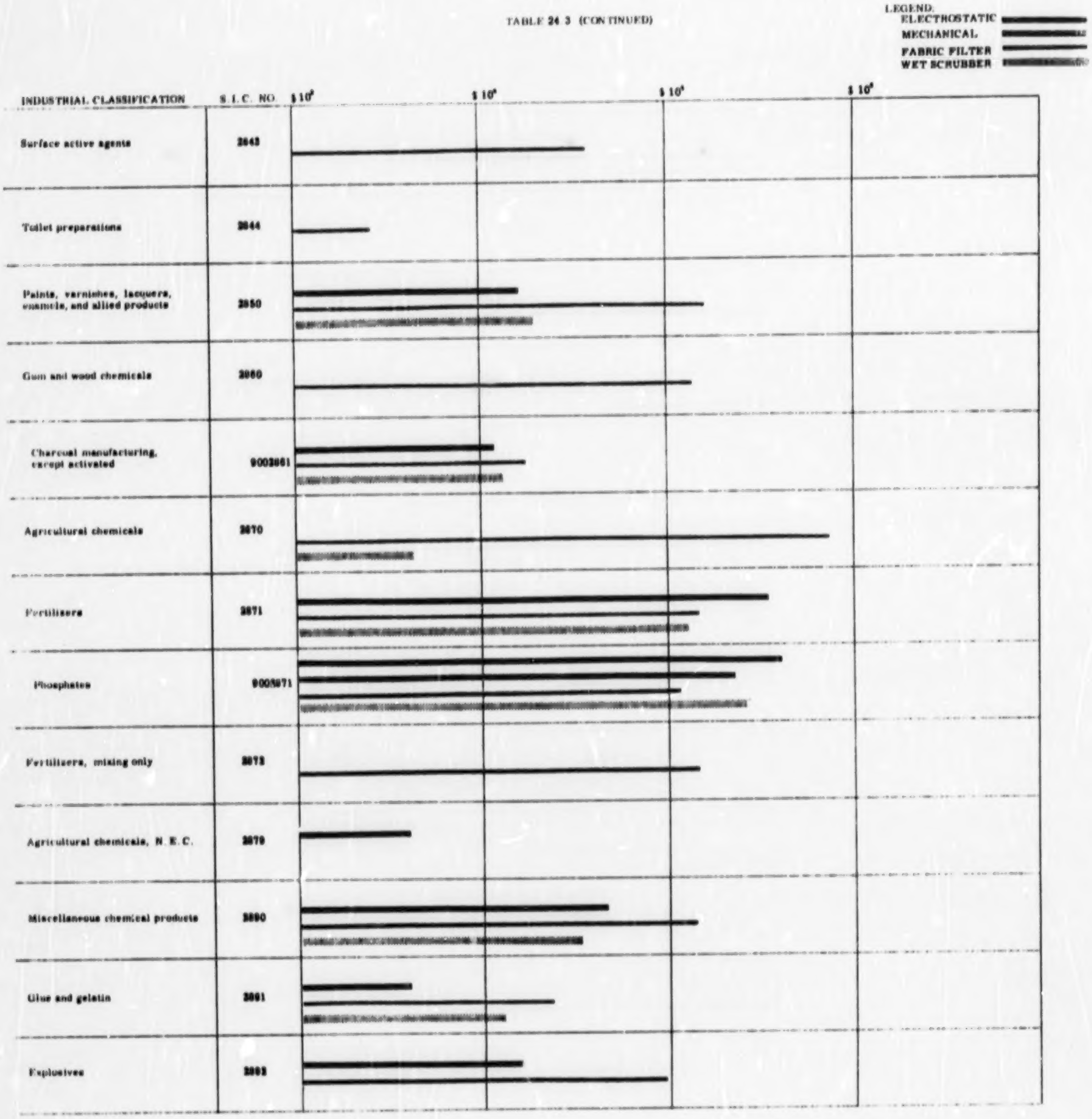

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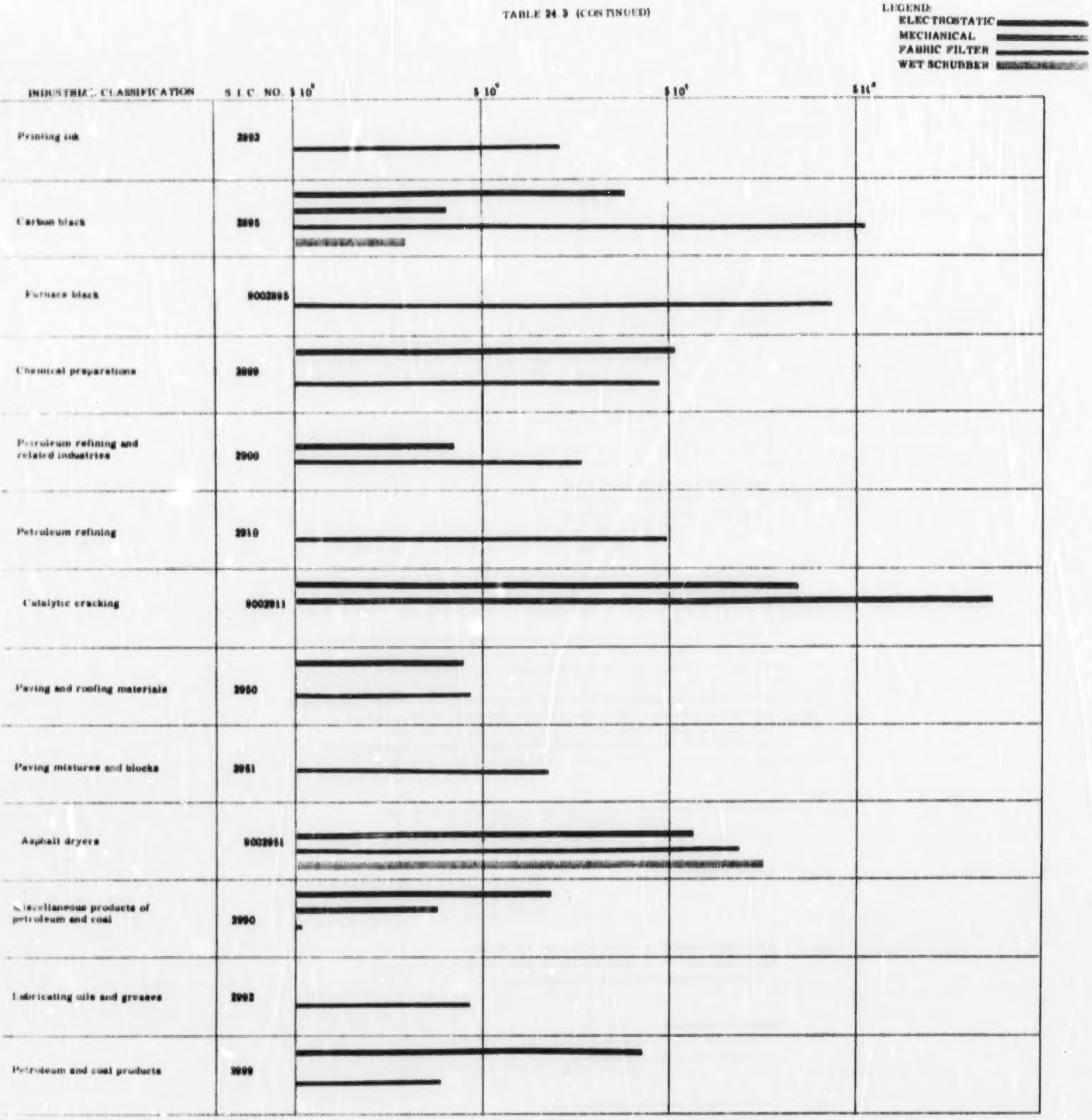

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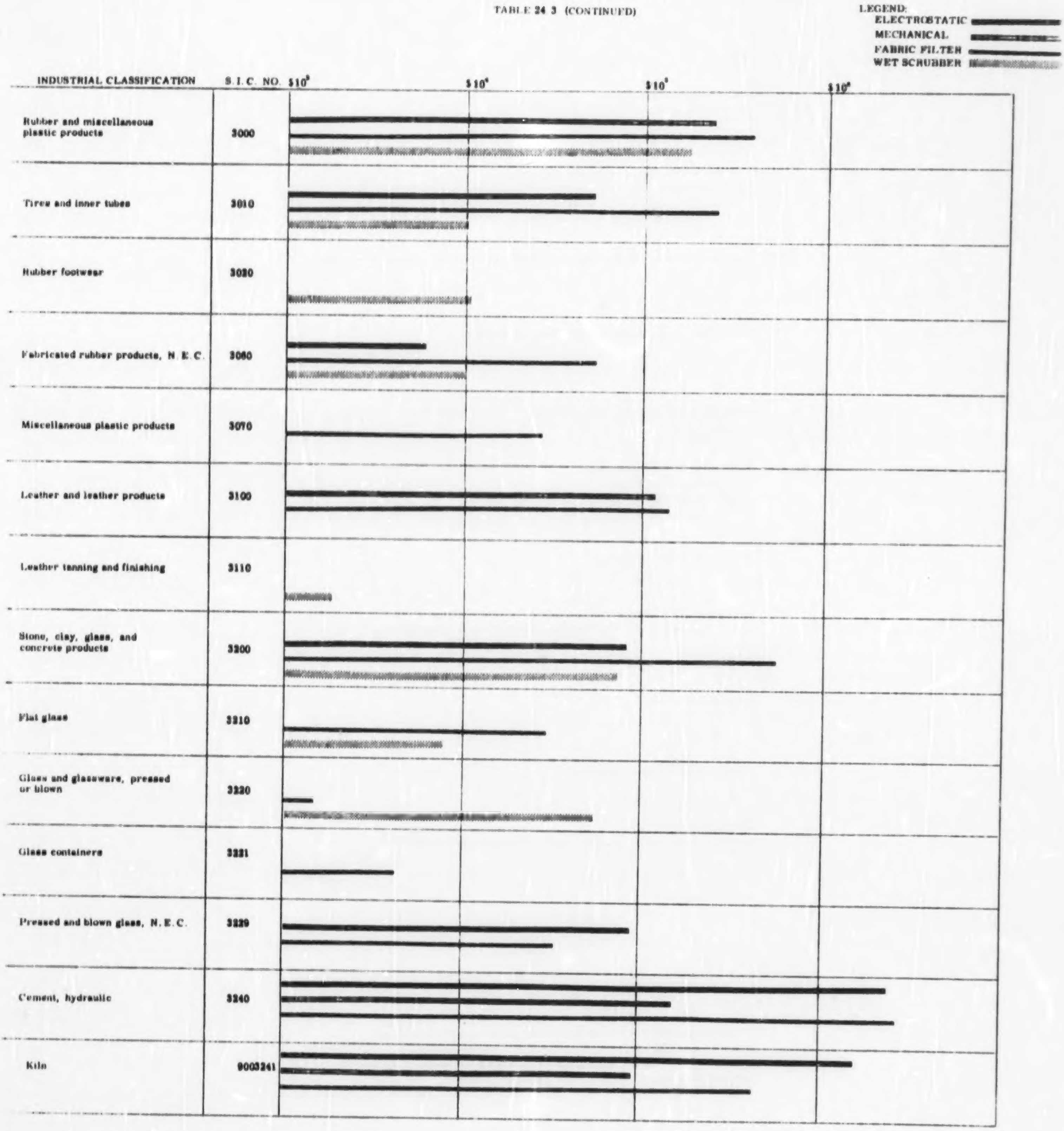

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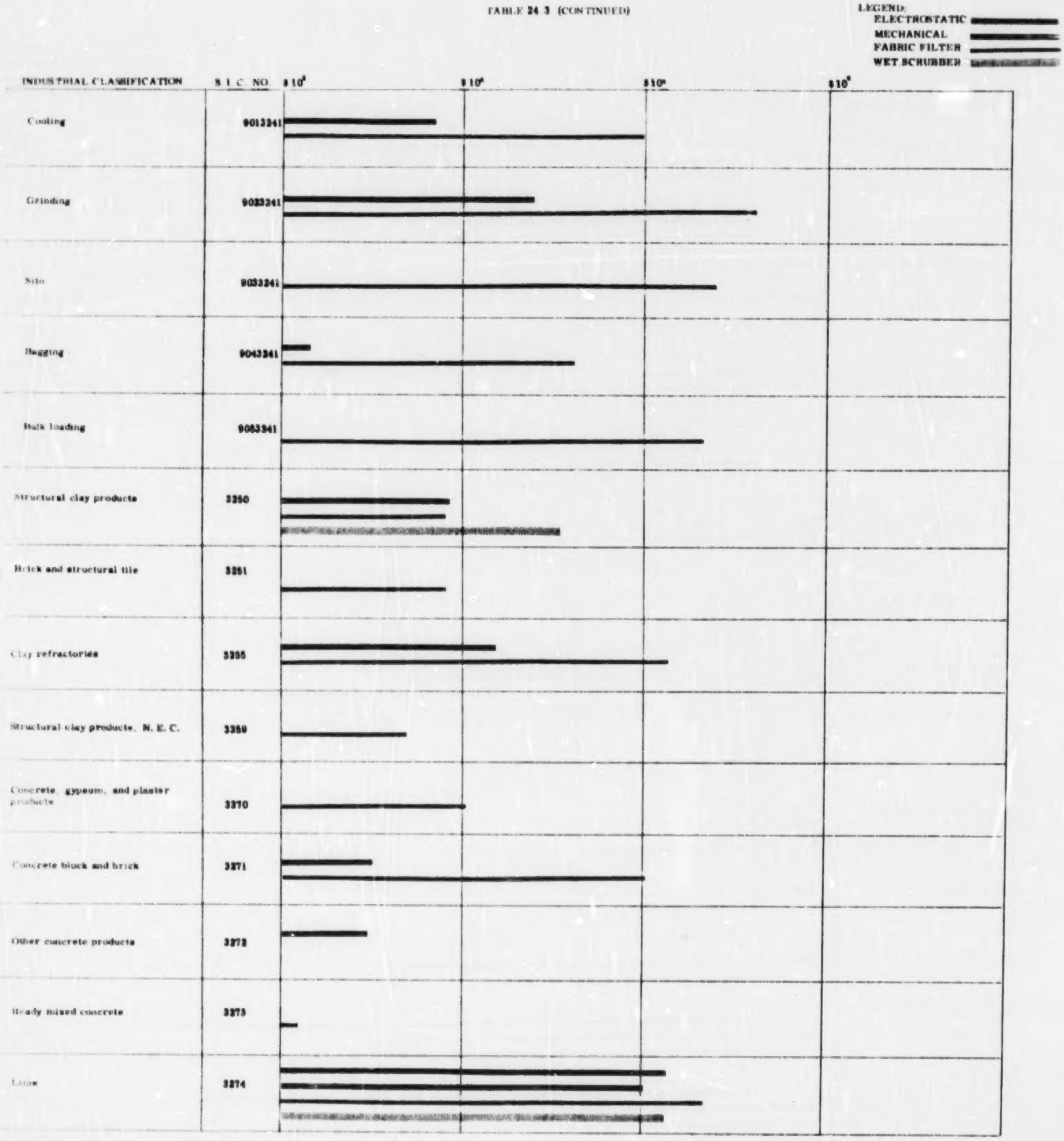

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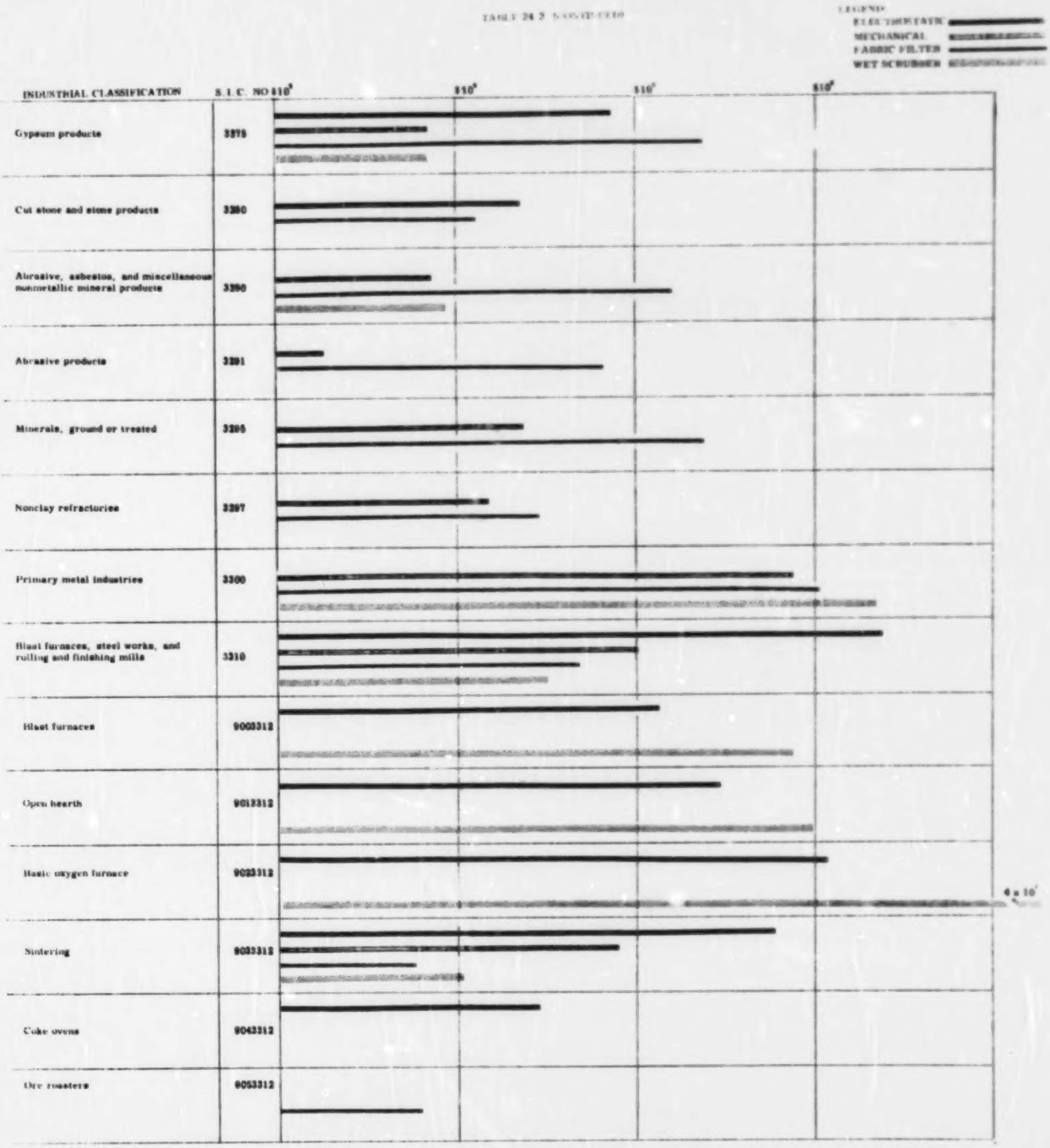

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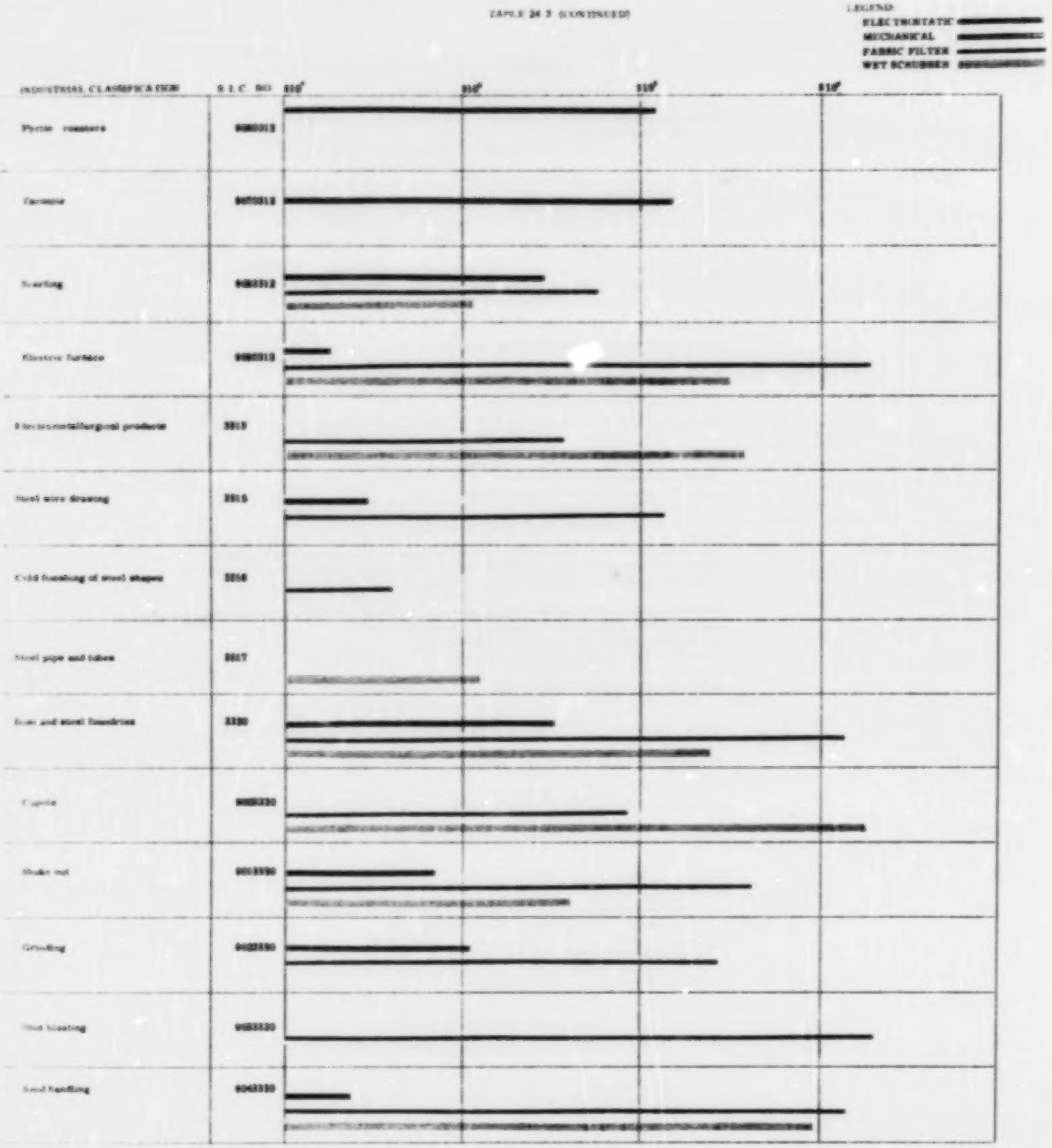
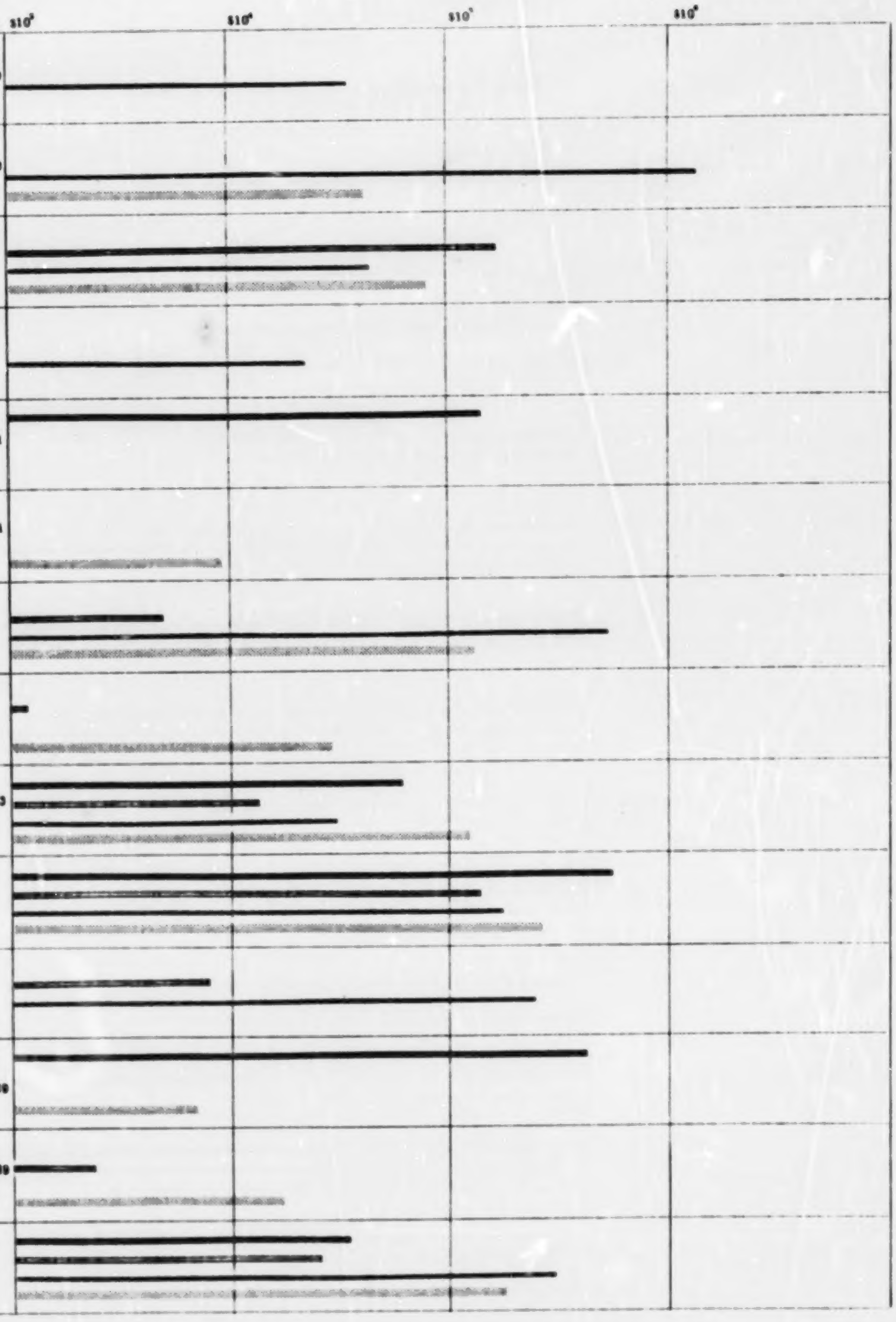


TABLE 24 3 (CONTINUED)

LEGEND
ELECTROSTATIC ▬
MECHANICAL ▬
FABRIC FILTER ▬
WET SCRUBBER ▨

INDUSTRIAL CLASSIFICATION	S.I.C. NO.
Electric induction furnace	9053320
Electric arc furnace	9083320
Primary smelting and refining of nonferrous metals	3330
Primary smelting and refining of copper	3331
Copper reverberatory furnace	9013331
Copper converter	9023331
Primary smelting and refining of lead	3332
Primary smelting and refining of zinc	3333
Zinc roaster	9003333
Primary production of aluminum	3334
Primary smelting and refining of nonferrous metals, N.E.C.	3339
Elemental phos.	9003339
Molybdenum	9033339
Secondary smelting and refining of nonferrous metals	3340



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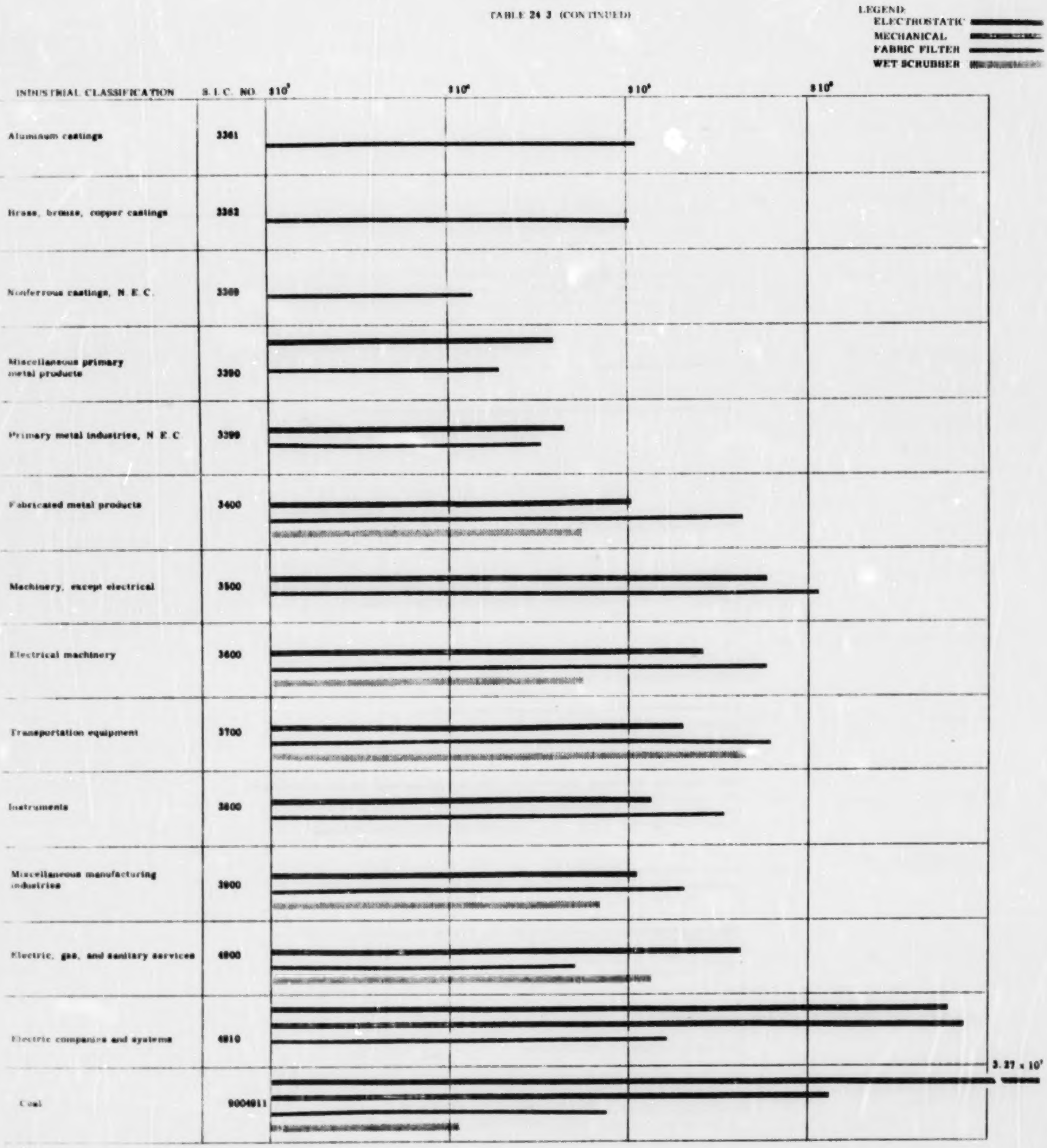

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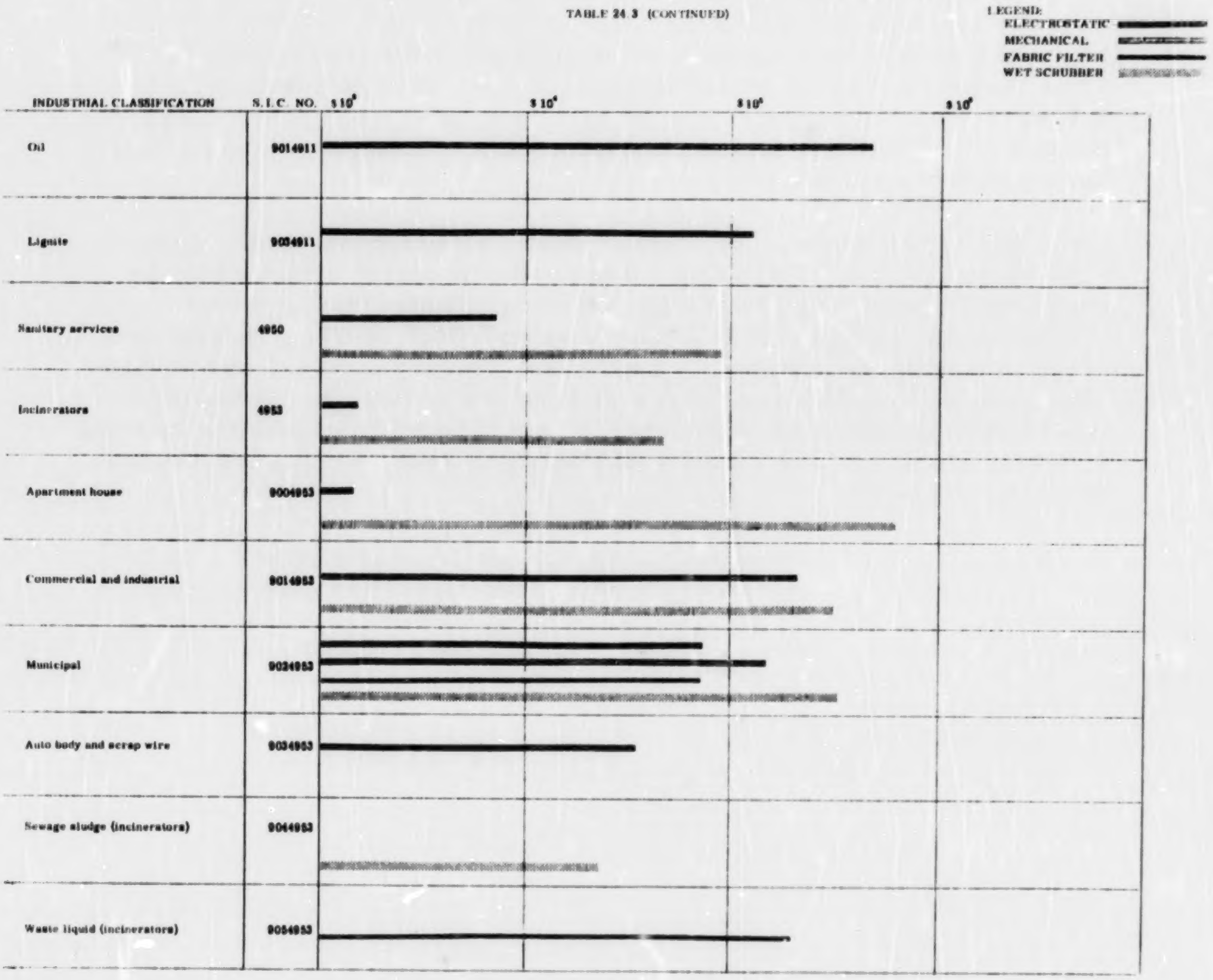

PARTICULATE CONTROL TOTAL:

ELECTROSTATIC	$57,695,531
MECHANICAL	22,028,931
FABRIC FILTER	37,709,161
WET SCRUBBER	23,753,477

One serious drawback in the use of precipitators for some dust collection problems has been an inability to obtain data on which to base a size determination if changes in the process or fuel are made. In some instances, precipitators designed to operate at efficiencies of 90% have resulted in efficiencies of 60-70% in practice. These instances occur with sufficient frequency that adverse attitudes toward precipitators are often developed.

The future application of electrostatic precipitators would appear to be along the lines of existing installations, probably enhanced by the necessity for high temperature gas cleaning. However, increased activity in air pollution control should extend electrostatic precipitator usage into areas of municipal incinerators, foundry cupolas, and perhaps other processes. Advances in precipitator technology, which would reduce precipitator size requirements, would further enhance the existing economic advantage associated with low power requirements for electrostatic precipitators.

CHAPTER 24
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